

✓ SIMULATION OF LONG TUBE VERTICAL EVAPORATOR

A Thesis Submitted
In Partial Fulfilment of the Requirements
for the Degree of
MASTER OF TECHNOLOGY

By
ASHOK KUMAR AGARWAL

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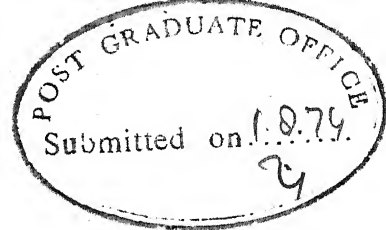
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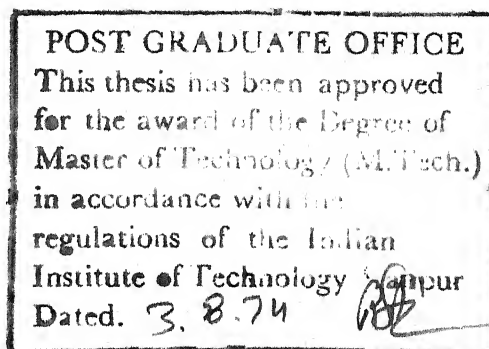
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CERTIFICATE

This is to certify that the present work, "Simulation of long tube vertical Evaporator" has been carried out under my supervision and has not been submitted elsewhere for a degree.

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ABSTRACT

✓ [A simulation programme is developed to predict the profiles of liquor temperature, concentration, heat flux, overall heat transfer co-efficient, pressure, pressure gradient, void fraction and vapor quality along the tube length of a natural circulation LTV evaporator. These quantities are predicted for the nonboiling and boiling sections of an evaporator tube from a combination of the following expressions: continuity, momentum and energy balance equations, and frictional pressure drop and holdup correlations for single phase and two phase flow, heat transfer correlation for the nonboiling, nucleate boiling and annular film regions, and correlations for the engineering properties of the process liquor such as thermal conductivity, viscosity, density, surface tension, boiling point rise and specific heat. A stepwise computation procedure has been developed starting with feed inlet conditions assuming a series of differential sections in the nonboiling and boiling sections of the tube and using the criteria of a maximum in liquor temperature profile and a minima in the pressure gradient curve for the transition from nonboiling to nucleate boiling and from nucleate boiling to annular film region respectively.

Results from this simulation programme agree with Badger's experimental data with an overall accuracy of 15% and agreement is within 3% for the temperature profile and an average deviation of 25% for the overall heat transfer co-efficient based on Gudmundson's experimental data for water. The model is applied to bamboo, baggase and pine black liquors,] ✓ over the concentration range 12-28% and the pronounced effect of viscosity

on heat transfer rates is illustrated for the following conditions:

Feed-rate 100 tons/hr, temperature 50°C and concentration 12.0%;

Evaporator-tube diameter 51 mm, length 7.6 m and number of tubes 300,

dirt scale resistance $R_D = 0.45 \text{ m}^2 \text{ }^\circ\text{K/KW}$. The nonboiling section length is 0.92 m and 2.13 m for bamboo and bagasse black liquors; the corresponding

values of the overall heat transfer coefficient are $2.4\text{--}2.7 \text{ KW/m}^2 \text{ }^\circ\text{K}$

(from 50°C, 0.9 cp to 106°C, 0.45 cp) and $0.8\text{--}1.1 \text{ KW/m}^2 \text{ }^\circ\text{K}$ (from 50°C,

4.7 cp to 102.4°C, 1.9 cp) with clean surfaces. In the boiling section

the overall heat transfer co-efficient values are $2.4\text{--}3.1 \text{ KW/m}^2 \text{ }^\circ\text{K}$ and

$2.5\text{--}3.3 \text{ KW/m}^2 \text{ }^\circ\text{K}$ for bamboo and bagasse black liquors respectively; the

final concentration, temperature and viscosity values are 27.8%, 95.3°C,

1.1 cp for bamboo and 25.9%, 86.2°C, 9.8 cp for bagasse black liquors

assuming clean tube surface. With the inclusion of the dirt scale

resistance the nonboiling length increases to 1.83 m and 2.75 m with a

corresponding decrease in heat transfer co-efficient to $1.2\text{--}1.3 \text{ KW/m}^2 \text{ }^\circ\text{K}$

and $0.6\text{--}0.7 \text{ KW/m}^2 \text{ }^\circ\text{K}$ for bamboo and bagasse black liquors respectively in

the nonboiling section. In the boiling region the dirt resistance

decreases heat transfer co-efficient to $1.2\text{--}1.4 \text{ KW/m}^2 \text{ }^\circ\text{K}$ and $1.2\text{--}1.45 \text{ KW/m}^2 \text{ }^\circ\text{K}$

for bamboo and bagasse black liquors respectively giving a final liquor

with concentration, temperature, viscosity of 15.7%, 103°C, 0.41 cp for

bamboo and 15.4%, 99°C, 2.0 cp for bagasse black liquor.

CHAPTER 1

INTRODUCTION

Natural circulation long tube vertical (LTV) evaporators are widely used for the concentration of solutions such as milk, fruit juice, spent pulping liquors etc. This study deals with the development of a model to predict changes in evaporation capacity and to recommend auxiliary heat exchange equipment that may be needed to accommodate a large change in the viscosity of the process liquor. A good example is the evaporation of weak black liquors from alkaline pulping process. Paper mills in India use bamboo as the major raw material and smaller amounts of mixed hardwoods, salai, baggase, grasses, pine etc. Increasing quantities of baggase and eucalyptus may also become available to the paper industry. The evaporation plant in a paper mill must be capable of handling bamboo, baggase, eucalyptus black liquors and also mixed black liquors whose viscosities vary considerably over the same range of concentration and temperature (38).

Fluid flow and heat transfer mechanisms in a LTV evaporator result in nonboiling, nucleate boiling and annular film regions along the tube length and determine the changes in solution temperature and concentration. In simulating the performance of a LTV evaporator process feed liquor conditions such as flow rate, temperature, concentration and pressure are usually available. Simulation consists in predicting the profiles of liquor temperature, concentration, heat flux and overall heat transfer co-efficient along the length of the evaporation tube. These computations would also give variation in pressure, pressure gradient, void fraction and vapor quality along the tube length. All these

quantities can be predicted for the given evaporator unit from a combination of the following expressions: Continuity, momentum and energy balance equations, and frictional pressure drop and holdup correlations for single phase and two phase flow, heat transfer correlation for the nonboiling, nucleate boiling and annular film regions, and correlations for the engineering properties of the process liquor such as thermal conductivity, viscosity, density, surface tension boiling point rise and specific heat. Computations would begin with the feed inlet conditions at the bottom of the evaporation tube and proceed through the nonboiling, nucleate boiling and annular film regions, and thus traverse the entire tube length. Computations are carried out in a stepwise manner assuming a series of differential sections of the evaporator tube and using the criteria of a maximum in liquor temperature profile and a minima in the pressure gradient curve for the transition from nonboiling to nucleate boiling and from nucleate boiling to annular film region respectively.

Simulation programme developed in this study is compared with the experimental results of Brooks and Badger⁽⁴⁾ and Gudmundson^(2,45) for evaporating water and subsequently used for the concentration of pine, bamboo and bagasse black liquors.

CHAPTER 2

Fluid Flow Pattern & Heat Transfer Mechanisms in a Long Tube Vertical Evaporator

LTV evaporators generally consist of 2.5 - 7.5 cm. diameter, 5-10 meter long tubes. Fluid flow patterns and heat transfer mechanisms obtainable for a liquor feed below its saturation temperature are shown in Fig.2.1. The following flow patterns and corresponding heat transfer mechanisms are normally observed:

1. Single Phase Liquid Flow (Convective Heat Transfer): Solution is heated to its saturation temperature by convection in the nonboiling section with the cold feed introduced at the bottom of the tube.
2. Bubble Flow (Nucleate Boiling): Superheating of the liquor adjacent to the tube wall occurs to some extent and small bubbles of vapor form on the heat transfer surface and results in a bubble flow pattern as they are released from the surface.
3. Slug Flow (Nucleate Boiling and Convective Heat Transfer): With an increase in number and size, the bubbles coalesce and two phase slug flow commences up the tube. In the beginning heat transfer takes place by nucleate boiling mechanism but with the increase in slug size, heat transfer takes place through liquor film by forced convection mechanism.
4. Annular Film Flow (Forced Convection): The increase in vapor velocity results in a central core of vapor flow and an annular climbing film of liquor and heat is transferred across this liquor film by forced convection.
5. Mist Flow (Forced Convection): With an increase in temperature gradient, dry patches begin to appear on the tube surface and liquid droplets are

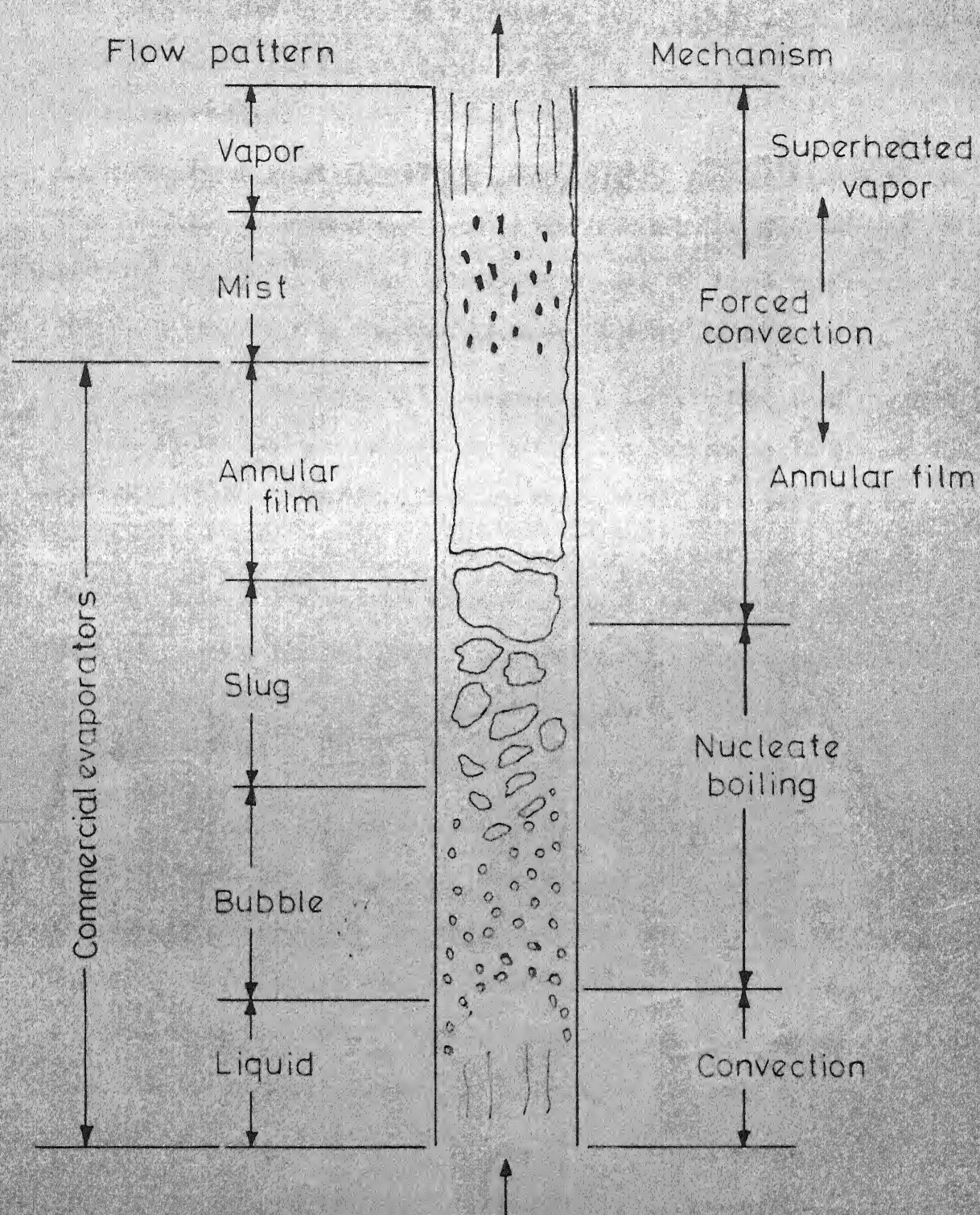


Fig.2.1- Fluid flow pattern and heat transfer in a LTV evaporator.

carried as a mist by the high velocity vapor. Heat transfer takes place through liquor film and at places of dry patches, directly to vapor by forced convection.

6. Single Phase Vapor Flow (Forced Convection): As the vapor rate is increased liquor entrainment becomes more and more predominant until, finally, the liquid annulus disappears and all the liquid is carried in the vapor core, giving essentially vapor flow up the tube.

In a commercial LTV evaporator, a certain length of the tube is used in heating the feed to the saturation temperature; in normal operation bubble flow, slug flow and annular film flow regions are encountered. Mist and single phase vapor flow patterns are not usual in chemical process evaporators unless very large temperature gradients are impressed on the system; further evaporation in a single pass is limited to less than 50-60%. Thus it is necessary to consider regions (1) to (4) listed above in a detailed analysis of heat transfer in a LTV evaporator.

The flow regions and the accompanying heat transfer mechanisms, shown in Figure 2.1 are not very distinct and there is considerable overlap in actual operation. For the purpose of this analysis the following heat transfer mechanisms are considered: 1) Convection (nonboiling section) 2) Nucleate boiling (bubble flow and part of slug flow region) 3) Forced convection (remainder of slug flow and annular film region).

Differences in fluid flow pattern and heat transfer mechanisms would result in a variation of the heat transfer co-efficient along the length of the evaporator tube. There is a slight maxima in liquor

temperature caused by pressure drop indicating the start of nucleate boiling region after the nonboiling section. Likewise transition from slug flow to the annular film flow regime is accompanied by a minima in the pressure gradient curve along the tube caused by the high void fraction of vapor (0.8-0.9) which in turn leads to high frictional pressure drop. This has been observed by Anderson et.al.⁽¹⁵⁾ in their experimental findings. These criteria may be used to predict approximately the sections of the tube lengths where single phase liquid flow, bubble & slug flow, and annular film flow regions occur during evaporation. Heat transfer co-efficients in these sections can be predicted using the appropriate available correlations. These are discussed in Chapter 4.

As the liquor traverses up the tube, liquor temperature increases due to sensible heat transfer and pressure decreases as a result of static and frictional forces in the nonboiling section; in this section liquor concentration remains constant as there is no evaporation. Changes in liquor temperature, pressure, pressure gradient and heat flux in a differential length of the nonboiling zone can be predicted by continuity, momentum & energy balance equations; and frictional pressure drop equation for the single phase liquid and heat transfer correlation. The transition to nucleate boiling occurs when the liquor temperature is equal to the saturation temperature corresponding to the pressure at that point in the tube. In the nucleate boiling region two phase flow starts with an increase in the volume of the vapor by evaporation accompanied by an increase in the liquor concentration. A differential length in this region can be considered for predicting the changes in liquor temperature and concentration, pressure,

pressure gradient, quality, void fraction and heat flux by applying continuity, momentum & energy balance equations alongwith void fraction, two phase frictional pressure drop and heat transfer correlations.

The stepwise procedure can be repeated until a minima appears in the pressure gradient along the length indicating the start of the annular film flow region. Changes occuring in this section can likewise be predicted by applying continuity, momentum & energy balance equations along with other equations to differential segments of this region in a stepwise manner until the evaporator tube length has been traversed. These stepwise calculation procedures are discussed in Chapter 5.

CHAPTER 3

ENERGY, MOMENTUM & CONTINUITY BALANCES FOR NONBOILING AND BOILING REGIONS

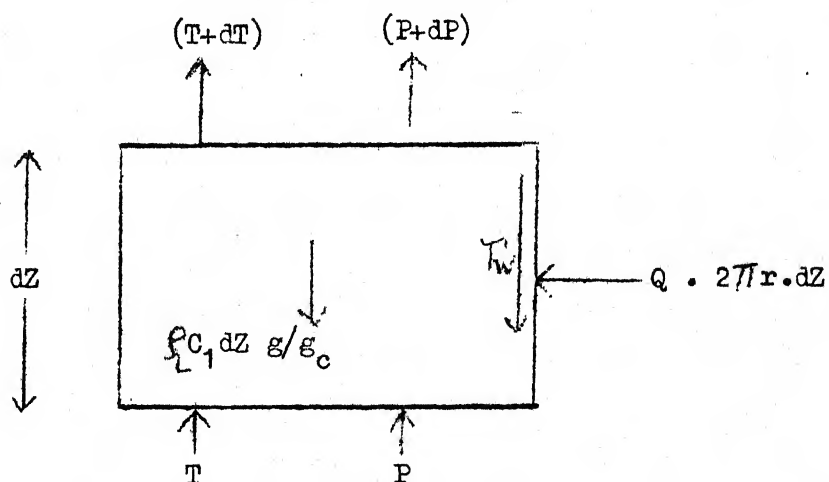
Energy, Momentum & Continuity balance equations are necessary for the simulation of the LTV evaporator. These equations will differ for the nonboiling section (single phase) and boiling section (two phase flow) of the evaporator tube length.

The appropriate balances applicable to the single phase region are discussed first. In the two phase flow regime energy, momentum & continuity balance expressions are obtained for (1) homogeneous model and (2) slip model. The former model assumes that liquor and vapor travel at same velocity where as the latter model is developed on the basis of different velocities for liquor and vapor phases.

A. SINGLE PHASE LIQUID

The following assumptions are made for the one-dimensional analysis of the single-phase flow existing in the nonboiling section of a

e



evaporator tube:

1. There is no radial variation of pressure, temperature, velocity & density.
2. Change in the Kinetic Energy is negligible.

Energy Balance:

Consider a differential section of the nonboiling section (dZ) shown above. Equation (3.1) gives the general differential form of the energy equation.

$$d(\text{Enthalpy}) + d(\text{Kinetic Energy}) + d(\text{Potential Energy}) = dq - dW \quad (3.1)$$

The terms in equation 3.1 can be evaluated separately as described below.

dH = Enthalpy of mixture leaving differential section

- Enthalpy of mixture entering differential section

$$= C_L dT, \text{ Btu/lb-m} \quad (3.2)$$

$$\text{Differential change in Kinetic Energy} = 0 \quad (3.3)$$

$$\text{Differential change in Potential Energy} = \frac{dZ}{J} \frac{g}{g_c} \quad (3.4)$$

The amount of heat transferred per pound mass of fluid flowing is given by equation 3.5.

$$dq = \frac{Q \cdot 2\pi r \cdot dZ}{W_L \cdot 3600} - \frac{Q \cdot 2\pi r \cdot dZ}{\int_L^P u \cdot C_1 \cdot 3600} \quad (3.5)$$

The amount of work done by the system on the surroundings

$$dW_s = 0 \quad (3.6)$$

Substituting equations 3.2 to 3.6 in equation 3.1, energy equation

gives:

$$dT = \frac{1}{C_L} \left[\frac{Q_c \cdot 2 \pi r \cdot dz}{C_1 u \rho_L \cdot 3600} - \frac{dz}{J} \frac{g}{g_c} \right] \quad (3.7)$$

Equation 3.7 can be used to predict the temperature change in a differential section of the nonboiling zone.

Momentum Balance:

Application of Newton's second law of motion to the fluid in the differential section (dz) gives the momentum balance equation 3.8.

$$\begin{aligned} \text{Forces of acceleration} - \text{forces of deceleration} \\ = \text{change in momentum of fluid} \end{aligned} \quad (3.8)$$

Change in the momentum of fluid is

$$\frac{\rho_L u C_1 (u + du)}{g_c} - \frac{\rho_L u C_1 u}{g_c} = 0 \quad (3.9)$$

When the equation 3.9 is stated mathematically, it becomes

$$C_1 P - (P + dP) C_1 - \frac{g}{g_c} \rho_L C_1 dz - \tau_w \pi D dz = 0 \quad (3.10)$$

Frictional pressure drop and shear stress at wall can be related

as

$$\tau_w = \frac{dP_F}{dz} \frac{C_1}{P} = \frac{dP_F}{dz} \frac{D}{4} \quad (3.11)$$

By using equation 3.11, equation 3.10 can be rewritten as:

$$dP = \frac{-g \rho_L dz}{g_c} - \left(\frac{dP_F}{dz} \right) dz \quad (3.12)$$

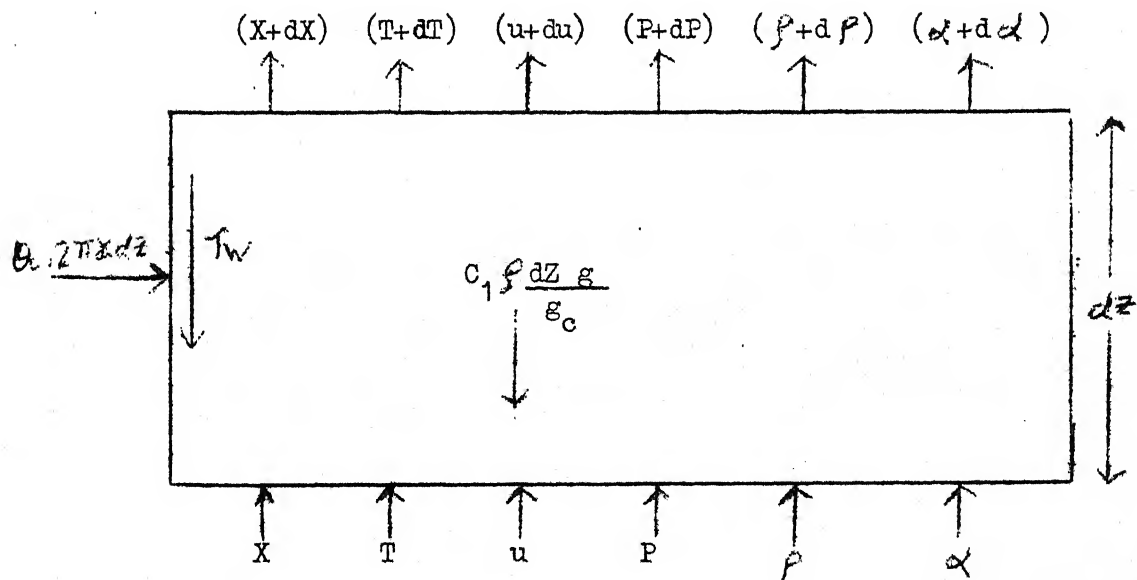
Equation 3.12 can predict the change in pressure along the tube length for the nonboiling section.

B. TWO PHASE FLOW (HOMOGENEOUS MODEL)

For making a one-dimensional analysis of the two-phase flow in the boiling section of an evaporator tube, the following assumptions are made:

1. No radial variation of pressure, temperature, velocity, quality and specific heat (This implies homogeneous vapor liquid mixture).
2. Relative velocity between the vapor and liquid is equal to zero. (no slip occurs between the vapor and liquid).
3. Vapor and liquid are in equilibrium.
4. Density of vapor-liquid mixture is given as

$$\rho = (1 - \alpha) \rho_L + \alpha \rho_V$$



Energy Balance:

Consider a differential section of the nonboiling section (dz) shown above. The general energy equation 3.1 is applied to this differential

section by evaluating each term separately.

$$\begin{aligned} dH &= \text{Enthalpy of mixture leaving differential section} \\ &- \text{Enthalpy of mixture entering differential section} \end{aligned} \quad (3.13)$$

The mixture leaving the section with a temperature $(T + dT)$ and a quality $(X + dX)$ had an enthalpy of

$$\begin{aligned} & (1-X-dX) C_L (T-T_O+dT) + (X+dX) \lambda + (X+dX) C_P (T+dT-T_O) \\ &= (1-X) C_L (T-T_O) + (1-X) C_L dT - (T-T_O) C_L dX + X \lambda \\ & \quad + \lambda dX + X C_P (T-T_O) + (T-T_O) C_P dX \end{aligned} \quad (3.14)$$

The mixture entering the section with a temperature T and a quality X had an enthalpy of

$$(1-X) C_L (T-T_O) + X \lambda + X C_P (T-T_O) \quad (3.15)$$

dH can be obtained from equations 3.14 and 3.15, and ignoring second order differentials gives equation 3.16.

$$dH = (1-X) C_L + \lambda dX + X C_P dT \quad (\text{if } T_O = T) \quad (3.16)$$

The differential change in Kinetic energy ($dK.E.$) is

$$\begin{aligned} d(K.E.) &= \frac{(u + du)^2}{2 g_c J} - \frac{u^2}{2 g_c J} \\ &= \frac{u du}{g_c J} \end{aligned} \quad (3.17)$$

The differential change in potential energy ($dP.E.$) is

$$d(P.E.) = \frac{dZ g}{J g_c} \quad (3.18)$$

The amount of heat transferred per pound mass of fluid flowing (dq) is

$$dq = \frac{Q \cdot 2 \pi r \cdot dZ}{3600 \cdot W_L} = \frac{Q \cdot 2 \pi r \cdot dZ}{f u C_1 \cdot 3600} \quad (3.19)$$

The amount of work (dW_s) done by the system on the surrounding is

$$dW_s = 0 \quad (3.20)$$

Combining equations 3.16 to 3.20 and equation 3.1 gives the energy equation for two-phase flow.

$$(1-X) C_L dT + \lambda dX + X C_P dT + \frac{u du}{g_c} + \frac{dz}{J} \frac{g}{g_c} = \frac{Q \cdot 2 \pi r \cdot dz}{\int u C_1 \cdot 3600} \quad (3.21)$$

Equation 3.21 can be used to predict temperature change or heat flux in the differential section.

Momentum Balance:

Momentum balance expression represented by equation 3.8 can be applied to the differential section to give equation 3.22.

$$\begin{aligned} C_1 P - (P + dP) C_1 - \int C_1 \frac{dz}{g_c} - \tau_w \pi D dz \\ = \frac{\int u C_1 (u+du)}{g_c} - \frac{\int u C_1 u}{g_c} \end{aligned} \quad (3.22)$$

By using equation 3.11, equation 3.22 can be rewritten for pressure drop in the differential section

$$-dP = \frac{\int dz}{g_c} + \left(\frac{dP_F}{dz} \right) dz + \frac{\int u du}{g_c} \quad (3.23)$$

Equation 3.23 can be used to predict pressure drop in the differential section.

Continuity Balance:

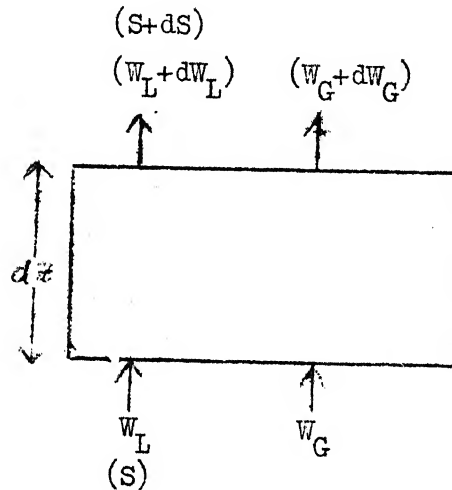
Continuity equation for the boiling section is represented by equation 3.24.

$$W_T = \int u C_1 \quad (3.24)$$

This can be given in differential form for steady flow conditions in the boiling section by equation 3.25

$$u \, dP + \int du = 0 \quad (3.25)$$

Liquor Solid Concentration:



Solids balance for the differential section is given by equation 3.26

$$\text{Total solid entering} = \text{Total solid leaving} \quad (3.26)$$

$$S W_L = S W_L + S dW_L + W_L dS + dW_L dS \quad (3.27)$$

Change in liquor concentration over the differential length of the evaporator is given by equation 3.28.

$$dS = \frac{-S \, dW_L}{W_L + dW_L} \quad (3.28)$$

Mass balance for the evaporation gives

$$dW_L + dW_G = 0 \quad (3.29)$$

Vapor quality of the mixture leaving the section is

$$X + dX = \frac{W_G + dW_G}{W_L + W_G} = X + \frac{dW_G}{W_L + W_G} \quad (3.30)$$

Equation 3.29 and 3.30 can be combined to give

$$dW_L = - (W_L + W_G) dX \quad (3.31)$$

By using equation 3.31 and 3.28, equation 3.32 can be obtained to predict the change in liquor concentration over a differential segment of the boiling section of the tube.

$$dS = \frac{S(W_L + W_G) dX}{W_L - (W_L + W_G) dX} \quad (3.32)$$

C. TWO PHASE FLOW (SLIP MODEL)

This model differs from the homogeneous model in that there is slip between liquid and vapor which is taken into account by the slip model. The following assumptions are made in the analysis.

1. No radial variation of pressure & temperature.
2. Vapor and liquid are in equilibrium.
3. Density of vapor-liquid mixture is given by

$$\rho = (1 - x) \rho_L + x \rho_v$$

Continuity Balance:

Equation 3.33 can be written for continuity balance

$$W_T = W_L + W_G \quad (3.33)$$

If the areas of cross section of the two streams are A_L and A_G .

$$W_L = \rho_L v_L A_L \quad (3.34)$$

$$W_G = \rho_G v_G A_G \quad (3.35)$$

The mass flux of each stream is then

$$G_L = P_L v_L (1 - \alpha) \quad (3.36)$$

$$G_G = P_G v_G \alpha \quad (3.37)$$

By using the definition of vapor quality (X), the above two equations can be used to give alternative expressions for the overall mass flux.

$$G = P_L v_L \frac{(1 - \alpha)}{(1 - X)} = \frac{P_G v_G \alpha}{X} \quad (3.38)$$

Momentum Balance:

Many alternative forms of the momentum equation can be derived by manipulating relationships among void fraction, vapor quality, total mass flux, liquor velocity, vapor velocity and other variables. For steady flow in a pipe Wallis⁽⁵³⁾ has given the following expression.

$$\frac{dP}{dZ} = \frac{-dP_F}{dZ} - \frac{G}{g_c} \frac{d}{dZ} \left[X v_L + (1 - X) v_G \right] + \left[\alpha f_G + (1 - \alpha) f_L \right] \quad (3.39)$$

Combining equation 3.39 with equation 3.38, momentum equation for the differential section can be written

$$\begin{aligned} \frac{dP}{dZ} = - \frac{dP_F}{dZ} = - \frac{G^2}{g_c} \frac{d}{dZ} \left[\frac{(1 - X)^2}{(1 - \alpha) P_L} + \frac{X^2}{P_G \alpha} \right] \\ - \left[(1 - \alpha) f_L + \alpha f_G \right] \end{aligned} \quad (3.40)$$

Equation 3.40 can be used to estimate pressure drop along the tube.

Energy Balance:

In energy balance, all terms will be alike as in the homogeneous model equation 3.21 except kinetic energy term

$$d(K.E) = \frac{1}{g_c J} d \left[\frac{X v_G^2}{2} + \frac{(1-X) v_L^2}{2} \right] \quad (3.41)$$

Combining equation 3.38 and 3.41, Kinetic energy term becomes.

$$\begin{aligned} d(K.E) &= \frac{1}{g_c J} d \left[\frac{G^2 X^3}{2 \rho_G^2} + \frac{(1-X)^3 G^2}{2 \rho_L^2 (1-X)^2} \right] \\ &= \frac{G^2}{2 g_c J} d \left[\frac{X^3}{\rho_G^2} + \frac{(1-X)^3}{\rho_L^2 (1-X)^2} \right] \end{aligned} \quad (3.42)$$

Substituting all terms in equation 3.1, energy equation becomes

$$\begin{aligned} (1-X) C_L dT + X C_p dT + \frac{G^2}{2 g_c J} d \left[\frac{X^3}{\rho_G^2} + \frac{(1-X)^3}{\rho_L^2 (1-X)^2} \right] \\ + \frac{dZ}{J} \frac{g}{g_c} = \frac{Q \cdot 2 \pi r \cdot dZ}{\rho u C_1 \cdot 3600} \end{aligned} \quad (3.43)$$

Equation 3.43 can be used for predicting temperature change in the differential section during two phase flow assuming slip model. The change in the liquor concentration for this model will be the same as for the homogeneous model and is given by equation 3.32.

Equation 3.1 to 3.12 are used for the nonboiling section and for the boiling section either equations 3.13 to 3.32 assuming homogeneous model for two phase flow or equations 3.32 to 3.43 based on slip model for two phase flow can be used.

REVIEW OF HEAT TRANSFER, PRESSURE DROP AND HOLD-UP CORRELATIONS
AND LIQUOR PROPERTIES

Fluid flow patterns and heat transfer mechanisms in a LTV evaporator were discussed earlier in Chapter 2. Changes occur in liquor conditions such as temperature, pressure, pressure gradient, concentration and factors relating to two-phase flow behavior such as quality and void fraction, and associated heat flux and overall heat transfer co-efficient along the length of the evaporator tube. All these quantities are determined in simulating the LTV evaporator. These can be predicted for a given evaporator unit from a combination of the following expressions: continuity, momentum & energy balance equations, and frictional pressure drop and holdup correlations for single phase and two phase flows, heat transfer correlations for the nonboiling, nucleate boiling and annular film regions, and correlations for the engineering properties of the process liquor such as thermal conductivity, viscosity, density, surface tension, boiling point rise and specific heat.

Momentum, energy and continuity relations for one dimensional single phase and two phase flow were discussed in detail in Chapter 3. This chapter deals with the following aspect 1) Critical review of heat transfer correlations for an LTV evaporator and recommended expressions for heat transfer, 2) frictional pressure drop and hold-up correlations to be used in this analysis and 3) Engineering properties for the process liquors.

A. Review of Heat Transfer in a LTV Evaporator:

include convective heat transfer, nucleate boiling and forced convective heat transfer. A summary of the available literature pertaining to these heat transfer mechanisms is given in Table 4.1.

Available heat transfer correlations are summarised separately for each of the three sections - non boiling, nucleate boiling and annular film regions. The complete correlating equations are not given; instead only the major variables that are included in the various investigations of heat transfer in LTV units are given. The tube size, liquor systems, accuracy of predictions and the range of application of these correlations are shown in Table 4.1 for the various expressions proposed by different investigators. It is necessary to choose expressions appropriate for the simulation of this investigation from amongst the various equations in Table 4.1. The use of empirical overall coefficients representing the cumulative effects of heat transfer in the different zones and the steam side characteristics are not suitable for the simulation purposes (Equations in section A, Table 4.1). The liquor flow in the non boiling section of a natural circulation LTV unit is in the laminar flow region for which the non boiling heat transfer correlations proposed by Badger et.al.⁽⁶⁾ and Martinelli & Bolter⁽⁴⁷⁾ can be used. Both of these expressions give essentially the same value of the non boiling heat transfer coefficient even though the latter include the effect of natural convection in their correlation (Appendix B). For computational purposes it is simpler to use Badger et.al.⁽⁶⁾ correlation than Martinelli & Bolter expression; in the latter expression non boiling

Table 4.1 : Summary of Correlations for Heat Transfer in a LTV Evaporator

Tube Size	Liquor System	Heat Transfer Correlation	Remarks
<u>A. Overall Heat Transfer Co-efficient (includes nonboiling and boiling regions)</u>			
<u>1. Uchida & Shinkawa (7)</u>			
L = 2.5 m	Water, H_2O-CH_3OH	$U = f(\lambda, \mu, V, F, Pr)$	Prediction within $\pm 15\%$
D = 11.9, 18.6, 25.5 mm			
<u>2. Gudmundson et.al. (2,45)</u>			
L = 10 m	Black Liquor	$U=f(p_1, p_2, p_3, p_4, p_5, C, C_c)$	A least square approach, within $\pm 15\%$ of original data.
D = 51/46 mm			
<u>B. Nonboiling Heat Transfer Co-efficient</u>			
<u>1. Sieder & Tate (46)</u>			
L = 1.55 m	Three different oils	$h=f(Re, Pr)$	Film heat transfer co-efficient. $Re > 2000$
D = 1.9 cm			
<u>2. Martinelli & Boelter (47)</u>			
L = 11.3 - 73.5 cm	Water, oil	$h=f(K, F_1, G^2, F_2, Gr, Pr, X, d_w)$	Film heat transfer co-efficient
D = 1.06-5.04 cm			Prediction within $\pm 20\%$ laminar range only
<u>3. Badger et.al (6)</u>			
L = 5.62 m	Water, sugar	$U_{NB}=f(C_p, \Delta T, f, \mu)$	Over all heat transfer co-efficient. Within $\pm 20\%$. $Re < 10,000$
D = 3.18 cm			

(Table 4.1 continued)

C. Boiling Heat Transfer Co-efficients (includes nucleate and annular film regions)

1.	<u>Brooks & Badger</u> (4)	Water	$U_B = f(W_f, \mu, \Delta t_{av}, V)$	Over all co-efficient. Prediction within $\pm 20\%$. Empirical in nature.
2.	<u>Stroebe & Badger</u> (5)	Water, sugar, Duponal solutions	$h_B = f(\nu, C_p, \mu, K, \nu, \Delta t_L)$	Film heat transfer co-efficient. within $\pm 20\%$. Empirical in nature.
3.	<u>Coulson & Mehta</u> (8)	Water	$h_B = f(\Delta t, W_f, \mu)$	Film co-efficient.
4.	<u>Pivet & Isbin</u> (9)	Water, Col ₄ , Isopropyl Alcohol, K ₂ CO ₃ solution	$h = f(d_w, K, U_m, \mu, C_p, \nu)$	Film heat transfer co-efficient. Empirical equation.
5.	<u>Gupta & Holland</u> (17, 18)	Hard water	$U_B = f(\Delta T, \nu, F, T_b, T_i, C_p)$	Over all heat transfer co-efficient within $\pm 16\%$

(Table 4.1 continued)

D. Nucleate Boiling Heat Transfer Co-efficient1. Mcnelly & Joulson (10)

L = 1.52 m Water, Isopropyl Alcohol, Ethyl Alcohol, Cyclohexane
 Various dia. tubes
 $Nu = f(d_w, Pr, Re_L, Re_v, \mu, \mu_L, P, \mu, f)$
 -low heat flux
 $Nu = f(Pr, q, d_w, \lambda, \mu_L, P, \mu, f)$
 -High heat flux

Film heat transfer co-efficients.

2. Guerrieri & Talty (48)

L = 1.97 m Cyclohexane, methanol, Benzene, pentene, heptane
 D = 2.54 cm
 $h = f(h_c, r_*, \zeta)$

Film heat transfer co-efficient.

3. Dengler & Addoms (11)

L = 6.07 m Water
 D = 2.54 cm.
 $h = f(h_L, F, X_{tt})$

Film heat transfer co-efficient

for forced circulation vaporization.

E. Annular Film Heat Transfer Co-efficient1. Guerrieri & Talty (48)

L = 1.97 m Cyclohexane, methanol, benzene, pentene, heptane
 D = 2.54 cm
 $h_c = f(h_L, X_{tt})$

Film heat transfer co-efficient

(Table 4.1 continued)

2.	<u>Denglar & Addoms</u> (11)	Water	$h = f(h_L, X_{tt})$	Film heat transfer co-efficient for forced circulation vaporization.
3.	<u>Anderson et.al.</u> (14)	Water	Set of equations	Film heat transfer-coefficient. Prediction within <u>+20%</u> . theoretical analysis.
4.	<u>Fennan & Tait</u> (16)	Water, $CH_3OH, C_2H_5OH,$ Chloroform, 2-propanol, Glycerol	$h = f(C_p, d_w, \rho_L, \sigma, \mu_V, \rho_V)$	Film heat transfer coefficient.
5.	<u>Kunze</u> (19)	Theoretical analysis	Set of equations	Film heat transfer coe-efficient. $10^{-3} < Pr < 10^4$

Nomenclature for this table are given below:

λ = liquid thermal conductivity

V = vaporization rate

F = feed rate

Γ = Mass rate of liquid per unit length of
periphery of the evaporator tube length.

$p1-p5$ = adjusted co-efficients

C	=	Correction factor	X	=	nonboiling length
$F1 \& F2$	=	Factors	ΔT	=	temperature difference
W_f	=	feed rate	m	=	factor
v	=	specific volume of vapor	Δt_L	=	liquid film temperature difference
U_m	=	log mean liquid vapor velocity	T_i	=	inlet temperature
T_b	=	boiling temperature	ψ	=	graphically read function of T
q	=	heat flux	P	=	pressure
h_c	=	annular film heat transfer coefficient	h_f	=	nucleate boiling heat transfer coefficient
h_L	=	single phase liquid heat transfer coefficient	r_*	=	radius of minimum size thermodynamically stable bubble
δ	=	laminar film thickness	X_{tt}	=	lockhart martinelli parameter
v_v	=	vapor velocity			
Gr	=	grashof No.			
Pe	=	peclet No.			
Gz	=	graetz No.			

section length is an implicit variable which will require a trial & error procedure for solution.

The heat transfer correlations given (Section C, Table 4.1) represent the boiling heat transfer co-efficient for the combined nucleate boiling and annular film regions and hence cannot be used to predict separately the nucleate boiling & annular film region heat transfer rates.

However, correlations have been proposed for the nucleate boiling region by several investigators (section D, Table 4.1). Guerrieri & Talty⁽⁴⁸⁾ correlation requires laminar film thickness which is often difficult to determine from operating units. Correlations proposed by Dengler & Addams⁽¹¹⁾ applies to forced circulation vaporization only. Mcnelly & Coulson⁽¹⁰⁾ expressions cover a wide range of equipment geometry & liquor systems and includes the effect of all major parameters. These authors have given two correlations for the nucleate boiling region applicable to low heat flux and high heat flux conditions, and can be used^{to} predict heat transfer coefficient.

Heat transfer correlations proposed by various authors for the annular film region are summarized in (Section E, Table 4.1). The prediction based on Guerrieri & Talty⁽⁴⁸⁾ correlation is rather poor and Dengler & Addams⁽¹¹⁾ correlation is applicable to forced circulation units only. Anderson et.al.⁽¹⁴⁾ theoretical analysis can be used but requires a rigorous computational procedure. Kunz⁽¹⁹⁾ theoretical analysis has not been substantiated by adequate experimental

data. Penman & Tait⁽¹⁶⁾ correlation is based on the use of two dimensionless groups and has been shown to fit very well with their experimental data and compare favourably with the predictions from Anderson et.al.⁽¹⁴⁾ theoretical approach. Hence correlation proposed by Penman & Tait can be used to simulate the heat transfer in the annular film region of a LTV evaporator tube.

A summary of the heat transfer correlations based on the above review for simulating heat transfer phenomena in a LTV evaporator tube is given below:

1. Convective Heat Transfer to Single Phase Liquid (Non boiling Region) -

Badger et.al.⁽⁶⁾

$$U_{NB} = 0.06 \left(\frac{C \Delta T_{NB} \rho^2}{\mu_{NB}} \right)^{0.7} \quad (4.1)$$

U_{NB} = Overall heat transfer coefficient for non boiling section, BTU/hr sq ft. °F

C = Specific heat of liquid, BTU/lb °F

ΔT_{NB} = Mean temperature difference for non boiling section, °F

ρ = Density of liquid, lb-m/cu.ft.

μ = Viscosity of the liquid lb/ft. hr.

2. Nucleate Boiling Heat Transfer Coefficient - Coulson & McNelly⁽¹⁰⁾

Low heat flux equation:

$$Nu = (1.3 + 39d) Pr_L^{0.9} Re_L^{0.23} Re_V^{0.34} \left(\frac{\rho_L}{\rho_V} \right)^{0.25} \left(\frac{\mu_V}{\mu_L} \right) \quad (4.2)$$

Where

$$\text{Nu} = \frac{h_B d}{k_L}, \quad \text{Pr}_L = \frac{C \mu_L}{k_L}$$

$$\text{Re}_L = \frac{4 W_f}{d \cdot \mu_L \cdot \pi}, \quad \text{Re}_V = \frac{Q \cdot L}{\lambda \cdot \mu_V} = \frac{4 W_V}{\pi d \mu_V}$$

High heat flux equation:

$$\frac{h_B d}{k_L} = 0.225 \left(\frac{C \mu_L}{k_L} \right)^{0.69} \left(\frac{Q \cdot d}{\lambda \cdot \mu_L} \right)^{0.69} \left(\frac{P \cdot d}{J} \right)^{0.31} \left(\frac{\rho_L}{\rho_V} - 1 \right)^{0.33} \quad (4.3)$$

d = Internal dia., ft.

h_B = Nucleate Boiling heat transfer co-efficient, BTU/hr.sq.ft.°F

k_L = Thermal conductivity, BTU/hr ft. °F

C = Specific heat, BTU/lb °F

μ = Viscosity, lb/ft. hr

W_f = Feed rate, lb/hr

Q = Heat flux, BTU/hr sq.ft.

L = Length of the tube, ft.

λ = Latent heat of vaporization, BTU/lb

P = Pressure, lb/sq ft.

ρ = Density, lb/Cu ft.

σ = Surface tension, lb/sq hr

3. Annular film heat transfer coefficient - Penman & Tait⁽¹⁶⁾

$$\frac{h}{C_{PL}} \sqrt{\frac{D}{\rho_L \sigma g_c}} = 0.012 \left[V_v \sqrt{\frac{D \rho_v}{g_c \sigma}} \right]^{0.5} \quad (4.4)$$

Where

h = film heat transfer co-efficient BTU/Sec. sq.ft. °F

C_{PL} = specific heat of liquor, BTU/lb °F

D = Diameter, ft.

ρ = Density, lb-m/cu ft.

σ = Surface tension, lb-f/ft.

V_v = Velocity of vapour, ft/sec.

g_c = Constant, 32.2 lb-m. ft/lbf. sec².

4. Shell - Side Heat Transfer Co-efficient-

The shell side heat transfer co-efficient for the filmwise condensation of steam can be determined from equation 4.5 (Meadams⁽⁵⁰⁾)

$$h = 1.13 \left[\frac{K_f^3 \rho_f^2 g \Delta H}{L \mu_f (T_{SV} - T_S)} \right]^{0.25} \quad (4.5)$$

Where

ΔH = Latent heat of condensation, BTU/lb

g = gravitational acceleration, ft/hr²

L = Length of tube, ft

T_S = Surface temperature, °F

T_{SV} = Saturated vapor temperature, °F

D = Outside diameter of tube, ft.

B. Frictional Pressure Drop & Hold-up Correlations:

Correlations representing frictional pressure drop and hold up are necessary for the prediction of pressure along the length of the evaporator tube length based upon energy, momentum and continuity balances.

Frictional pressure drop in the single phase liquid can be estimated from equations 4.6 - 4.11, given by Knudson & Katz⁽³¹⁾.

$$\Delta P_f = \frac{4f}{D} \frac{LV^2}{2 g_c} \quad (4.6)$$

The effect of roughness characteristics of the tube on the value of friction factor can be determined from equations 4.7, 4.8 and 4.9.

For laminar flow : $f = 16/N_{Re}$ (4.7)

For turbulent flow: $\frac{1}{\sqrt{f}} = 4 \log \frac{d_w}{e} + 2.28$ (4.8)

for $\frac{r_w/e}{R_e \sqrt{5}} > 0.005$

For transitation flow:

$$\frac{1}{\sqrt{f}} = 4 \log \frac{d_w}{e} + 2.28 - 4 \log \left[1 + 4.67 \frac{d_w/e}{R_e \sqrt{f}} \right] \quad (4.9)$$

For the case of nonisothermal flow in pipes, there is small change in the value of friction factor as given by Perry⁽²⁹⁾ and Seider et.al.⁽³⁰⁾. In the case of heating

Divide f by $(\mu_a/\mu_w)^{0.38}$ for laminar region (4.10)

Divide f by $(\mu_a/\mu_w)^{0.17}$ for turbulent region (4.11)

(The units to use in equations 4.6 to 4.11 are as follows:

- D = Tube Dia., ft
 L = Tube length, ft
 ρ = Fluid density, lb/Cu ft
 V = Fluid velocity, ft/sec.
 f = Friction factor
 ΔP_f = Pressure drop due to friction lb_f/sq ft.
 d_w = Tube dia. inch
 r_w = Radius of tube, inch
 μ_a = Viscosity at mean temp., cp
 μ_w = Viscosity at wall temp., cp)

The prediction of pressure drop in the boiling section would involve a detailed consideration of two-phase liquid-vapor flow patterns. Frictional pressure drop & hold-up correlation for the two phase flow have been reviewed by several authors^(21,24,33,34,52-61) and have recommended the use of Lochart & Martinelli⁽²⁴⁾ correlation for two phase frictional pressure drop given by equation 4.12.

$$\left(\frac{\Delta P}{\Delta L} \right)_{\text{fric}} = \left[\frac{2 \left(\frac{4}{\pi} \right)^{2-n} C_1 \mu_1^{1/4} W_1^{2-n}}{D_p^{5-n} \rho_1 g} \right] X^{n-2} \left(\frac{D_p}{D_1} \right)^{5-n} \quad (4.12)$$

The calculation of two phase flow pressure drop by Lochart & Martinelli analysis is discussed in Appendix A.

Levy⁽⁶²⁾ correlation can be used to predict the hold-up for

the flow of air water & liquid-vapor mixtures in vertical pipes.

$$\frac{(1-X)^2}{1-\alpha} + \frac{X^2 \rho_L}{\alpha \rho_G} - \frac{(1-X)^2}{2(1-\alpha)^2} = 0 \quad (4.13)$$

Where X = Quality of vapor

α = Vapor volume fraction

ρ_L = Density of liquid

ρ_G = Density of vapor

C. Engineering Properties of Process Liquors:

The use of momentum, continuity and energy, frictional pressure drop, hold-up & heat transfer correlation for the simulation of long tube vertical evaporator requires a compilation of the engineering properties and characteristics of commercial process liquors. These properties include density, viscosity, specific heat, thermal conductivity, surface tension, boiling point elevation and vapor pressure data. This study deals with the evaporation of black liquors from alkaline pulping processes. Paper industries in India use bamboo as the main cellulosic raw material. Other fibrous raw materials used in current practice include mixed hard-woods, baggase, grasses, straws, salai and pinewood; the use of these raw materials is rather limited and is determined by their availability to the paper mills. It is hoped that in the next four-five years the use of eucalyptus will be gradually established as a important pulping raw material. Several mills use bamboo as the major raw material and meet the balance of their requirements from same or the other cellulosic raw materials listed above. These raw materials are normally pulped separately owing to diverse differences in their physical,

chemical and morphological properties. The spent liquors from these pulping operations are usually mixed and treated for chemical recovery. The weak black liquor contains 12-22% dissolved solids; approximately two-thirds of the black liquor solids are organic compounds dissolved from the fibrous raw material during alkaline pulping reactions. The remaining one third portion of the black liquor solids constitute the spent inorganic chemicals. The organic compounds include lignin and carbohydrate degradation products, and are dissolved in black liquor as their sodium salts. Other organic compounds include soaps (sodium salt of resin and fatty acids present in the wood). This often leads to foam problems during handling of weak black liquors. The inorganic compounds in the black liquor include residual sodium hydroxide, sodium sulfide, sodium carbonate and smaller amounts of sodium sulfate, sodium thiosulfate, sodium sulfite etc. Black liquors from the pulping of bamboo, bagasse and agricultural residues usually contain a appreciable concentration of silica as sodium silicate in solution. This usually leads to scaling problems in chemical recovery equipments requiring frequent shutdowns for maintenance, when the weak black liquor is concentrated to 45-50% solid concentration in multiple LTV evaporators consisting of 5-7 effects. The concentrated black liquor is subsequently burnt in the recovery boilers for reclaiming pulping chemicals and to utilize calorific value of the organic compounds in the black liquor solids.

Even though bamboo is a major raw material in use today, it is possible that substantial quantities of bagasse from sugar mill and eucalyptus from plantation may become available to the paper industry. Such a situation would then require that a given pulp mill be in a position

to handle two or more major fibrous raw materials. The chemical recovery units must be capable of handling the black liquors from different fibrous sources. Koorse⁽³⁸⁾ has demonstrated the diverge differences in some of the engineering properties specially viscosity of bamboo, baggase and eucalyptus black liquors.

We can foresee a condition in the next few years when a particular paper mill will be handling bamboo for five to six months and baggase for the rest of the year. It would be possible to predict the changes in the evaporation plant on switching from bamboo to baggase using the simulation programme of this investigation. This would however require numerous engineering data for bamboo and baggase black liquors. The latter information has been experimentally obtained using commerical bamboo and baggase black liquors from Indian paper mills by Koorse⁽³⁸⁾. Correlations for the engineering properties of the liquors are summarized below:

1. Properties of Bamboo Black Liquor - Koorse⁽³⁸⁾

Viscosity: $\mu = \text{Exp} (A + B.S + C.S^2 + D.S^3)$, cp (4.14)

Where

$$A = 4.4014 - 0.2248 T + 0.0018 T^2$$

$$B = -0.479 + 0.0296 T - 0.00025 T^2$$

$$C = 0.02075 - 0.0013 T + 0.00001 T^2$$

$$D = -0.000175 + 0.00015 T - 0.0000001 T^2$$

Density: $\rho_L = 1.0182 + 0.00779 S - (0.000219 + 0.000003 S) \cdot T$, g/c (4.15)

Thermal Conductivity: $K = 0.53354269 - 0.003425 S + (0.0011355 + 0.00000209 S)T$
KCal/hr. m°K (4.16)

$$\text{Boiling point Elevation: B.P.E.} = 0.19 S, ^\circ\text{C} \quad (4.17)$$

Equation 4.17 for boiling point rise was obtained from limited experimental data and hence needs further refinement with additional data. Nevertheless it can be used as an approximate expression to include the effect of boiling point rise in the evaporator calculations.

2. Properties of Bagasse Black Liquor - Koorse⁽³⁸⁾

$$\text{Viscosity: } \mu = \text{Exp} (A + B.S + C.S^2), \text{ cp} \quad (4.18)$$

$$\text{where } A = 18.276186 - 0.507619 T + 0.00343 T^2$$

$$B = -1.427423 + 0.043054 T - 0.0003074 T^2$$

$$C = 0.032877 - 0.00898 T + 0.0000064 T^2$$

$$\text{Density: } \rho_L = 1.0207 + 0.0071 S - T (0.000279 + 0.0000012 S) \text{ g/cc} \quad (4.19)$$

$$\begin{aligned} \text{Thermal Conductivity: } K &= 0.050579 - 0.002563 S + (0.000899 + 0.00000749) S \\ &\text{KCal/hr.m}^\circ\text{K} \quad (4.20) \end{aligned}$$

$$\text{Specific Heat: } C_L = 0.993 - 0.01258 S + (0.0000556 + 0.000104 S) \quad (4.21)$$

$$\text{Boiling Point Elevation: B.P.E.} = 0.14 S, ^\circ\text{C} \quad (4.22)$$

Equation 4.22 can be used as an approximate expression to include the effect of boiling point rise in the evaporator calculations.

3. Properties of Pinewood Black Liquor - Gudmundson⁽³⁵⁾

$$\text{Viscosity: } \mu = \text{Exp} (A + B.S + C.S^2 + D.S^3), \text{ cp} \quad (4.23)$$

$$\text{Where } A = 0.4717 - 0.02472 T + 0.7059 \times 10^{-5} T^2$$

$$B = 0.06973 - 0.5452 \times 10^{-3} T + 0.1656 \times 10^{-5} T^2$$

$$C = 0.002046 + 0.3183 \times 10^{-3} T - 0.9761 \times 10^{-7} T^2$$

$$D = 0.5793 \times 10^{-4} - 0.6129 \times 10^{-6} T + 0.1837 \times 10^{-8} T^2$$

$$\text{Density: } \rho_L = 1007 - 0.495 T + 6.0 S, \text{ Kg/m}^3 \quad (4.24)$$

$$\text{Specific Heat: } C_L = (1.0 - 0.0054 S) 0.4187, \text{ WS/Kg } ^\circ\text{K} \quad (4.25)$$

4. Properties of Water -

$$\text{Viscosity: } \mu = 1.996531 - .0160284 T + .00006745 T^2, \text{ cp} \quad (4.26)$$

$$\text{Thermal Conductivity: } K = 587(1 + 0.00281 (T - 20)) 2.389 \times 10^{-6} \text{ Cal/cm.sec. } ^\circ\text{C} \quad (4.27)$$

5. Vapor Pressure-Temperature Relationship -

Vapor pressure and temperature can be correlated by Antoine equation 4.26 or 4.27

$$\log_{10} P = A - \frac{B}{T - \text{B.P.E.} + C} \quad (4.28)$$

$$T = \frac{B}{A - \log_{10} P} - C + \text{B.P.E.} \quad (4.29)$$

Values of Antoine equation constants are given in Table C-1 & C-2 of Appendix C respectively for equation 4.26 and 4.27.

(The units to use in equations 4.14 to 4.27 are as follows

P = Pressure, mm of Hg.

T = Temperature, $^\circ\text{C}$

S = Solid concentration, wt %)

CHAPTER 5

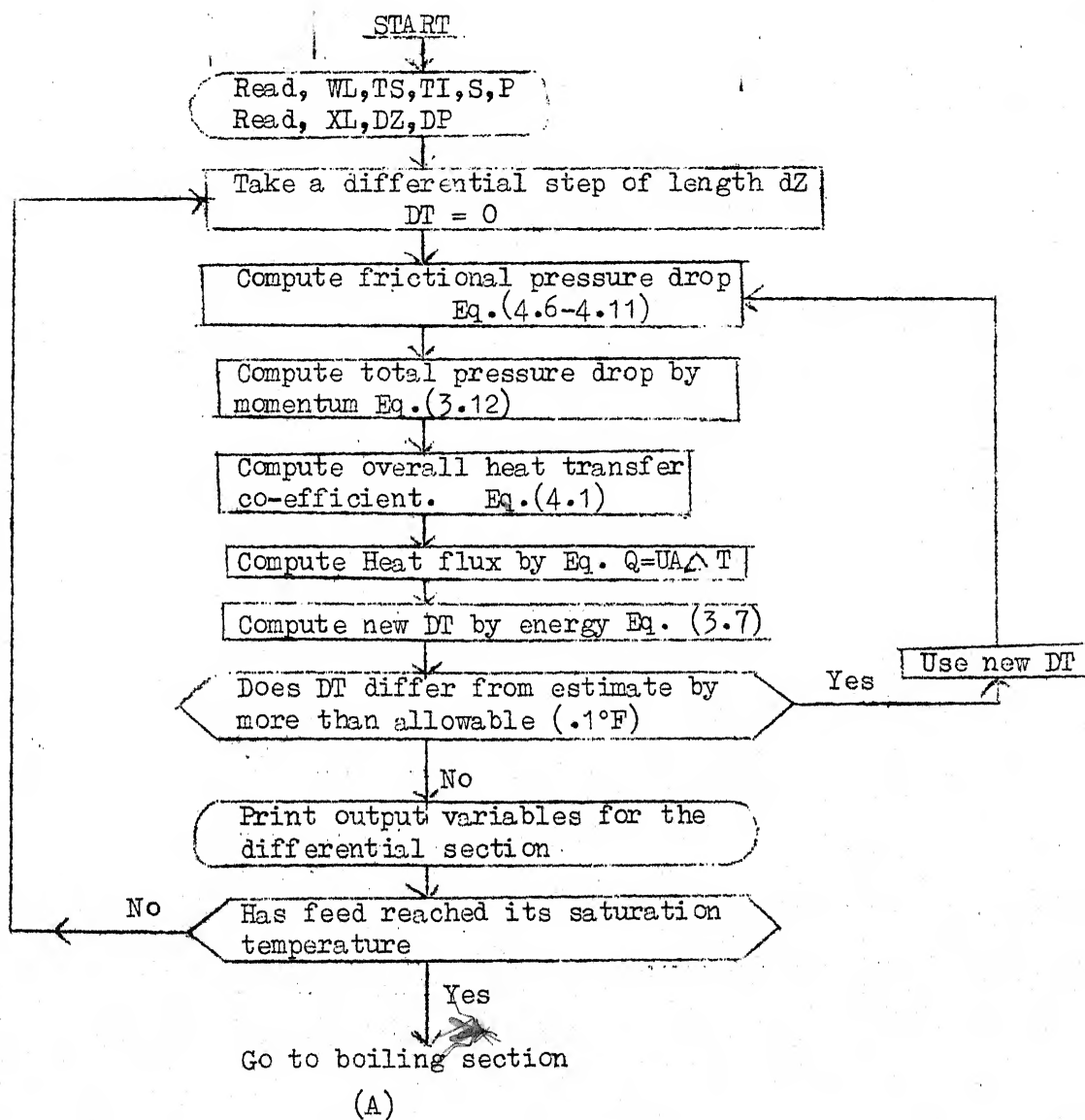
COMPUTATIONAL SCHEME FOR LONG TUBE VERTICAL EVAPORATOR

In simulating the performance of a LTV evaporator, process feed liquor conditions such as liquor flow rate, concentration of solids, temperature and pressures are usually available. Simulation consists in predicting the profiles of liquor temperature, concentration, heat flux and overall heat transfer co-efficient along the length of the evaporator tube. These computations would also give variation in pressure, pressure gradient, void fraction and vapor quality along the tube length. Computations would begin with the feed inlet conditions at the bottom of the evaporator tube and proceed through the nonboiling, nucleate boiling and annular film regions, and thus traverse the entire tube length.

Simulation computations are carried out in a stepwise manner assuming a series of differential sections of the evaporator tube and using the criteria of a maxima in liquor temperature profile and a minima in the pressure gradient curve for the transition from nonboiling to nucleate boiling and from nucleate boiling to annular film region respectively.

For a weak black liquor feed below its saturation temperature, flow chart of computation for the nonboiling section is given in Figure 5.1. Flow rate, concentration, temperature and pressure of feed liquor, steam chest temperature and tube geometry are fixed for starting the calculations. A differential length dZ of tube (.15 meter) is considered to determine the heat transfer rates. The liquor temperature in this differential section is assumed to be isothermal at first. The change in pressure in this section

Fig. 5.1 Flow Diagram for Nonboiling Section of a LTV Tube



can be calculated from the frictional pressure drop correlation equations (4.6 - 4.11) and momentum balance equation (3.12). The overall heat transfer co-efficient for nonboiling section can be obtained by Badger correlation equation 4.1. Heat flux can be computed by the heat transfer rate equation ($Q = UA\Delta T$). The next step consists of determining differential change in the temperature by energy equation 3.7. This new value of DT is then compared with the previously assumed value for the start of the computations. If they differ by more than a certain specified limit (.1 °F), then the calculations are repeated with this new value of DT . In case the difference is within specified limit then the calculated liquor temperature is compared with the saturation temperature corresponding to the estimated pressure at the exit of the differential section. If the liquor temperature happens to be below the saturation temperature, the calculations are repeated for the next differential section. This iterative procedure is repeated for several differential sections until the liquor temperature equals the saturation temperature of liquor. This point corresponds to the end of the nonboiling regime and subsequent calculations are carried out for the boiling section.

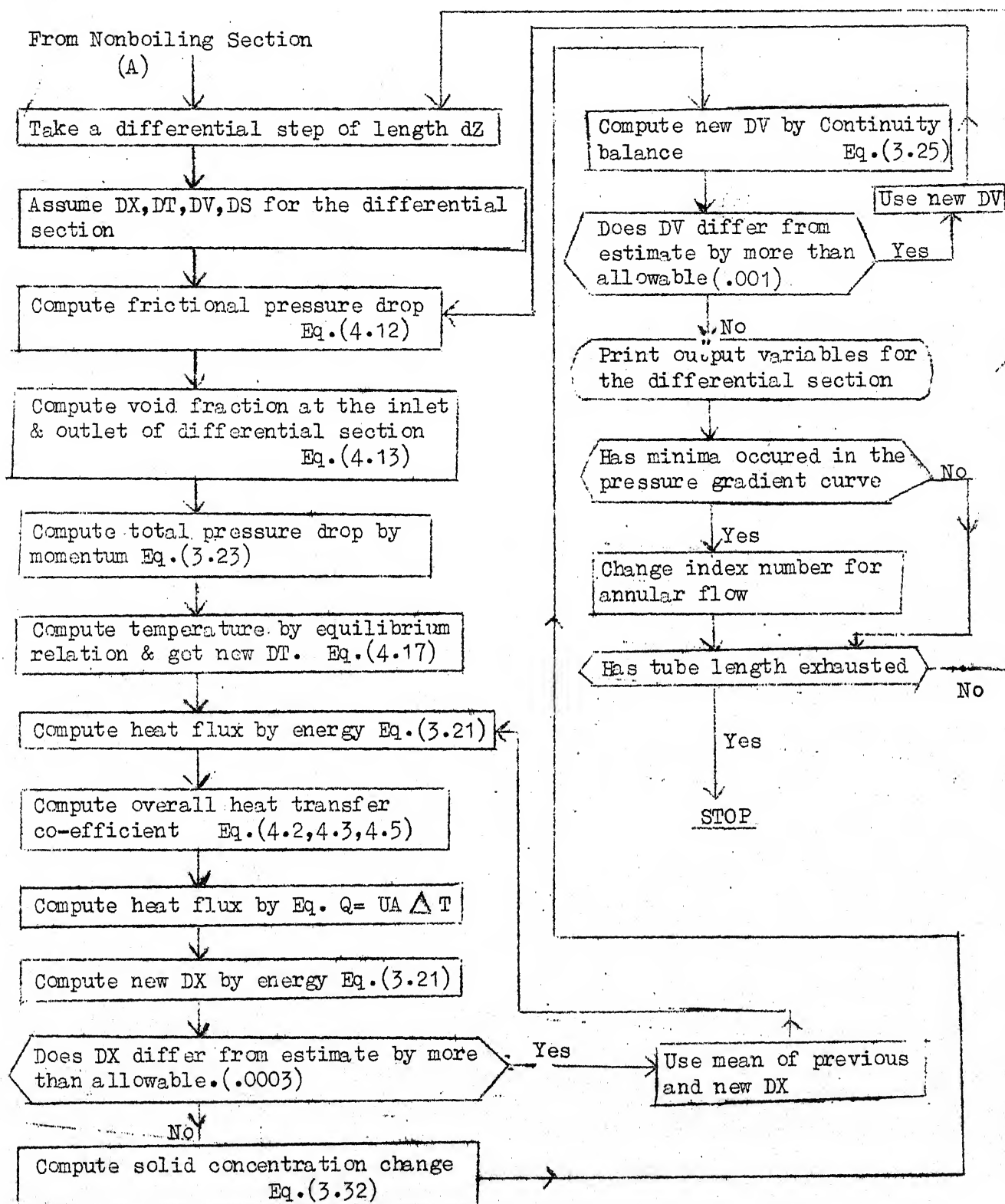
For the boiling section, simulation computations can be performed by assuming either homogeneous model or slip model for the two phase vapor-liquid mixture flow as given below:

1. Homogeneous Model:

Homogeneous model assumes that the velocity of liquid and vapor are same and there is no slip between liquid and vapor. The

computational scheme for estimating changes in the process liquor conditions and heat transfer rate is shown in Fig. 5.2 for the boiling section. Conditions of the liquor feed to the boiling section would be the same as exit liquor conditions from the nonboiling section. Now consider a differential length dZ (0.15 meter) in the evaporator tube length. The calculation for this differential section must give the change in liquor temperature, pressure, pressure gradient, concentration, quality, void fraction, heat flux and overall heat transfer co-efficients. Firstly values for the change in the quality, temperature, solid concentration and velocity (DX, DT, DS, DV respectively) for this differential segment are assumed. Two phase flow frictional pressure drop is calculated by Lockhart and Martinelli correlation (equation 4.12). The void fraction at the inlet and exit of the differential zone can be calculated by Levy correlation (equation 4.13). Obviously the void fraction at the entrance to the boiling section will ^{be} zero and gradually increases as boiling progresses along the evaporator tube. The total pressure drop can be obtained by momentum balance (eq. 3.23). The temperature corresponding to the calculated pressure can be determined from the Antoine eq. 4.17 and this would give new value of change in temperature for the differential section. Heat flux for the differential section can be obtained by the energy equation 3.21. Overall heat transfer co-efficients can be calculated by using eq. 4.2, 4.3 and 4.5. New value of heat flux can be calculated by heat transfer rate equation ($Q = UA \Delta T$). This value of heat flux can be used in energy eq. 3.21 to calculate new value of change in quality for the differential section and is compared with the previously assumed value of DX . If the difference between these two values of DX is

Fig. 5.2 Flow Diagram for Boiling Section (Homogeneous Model) of a LTV Tube



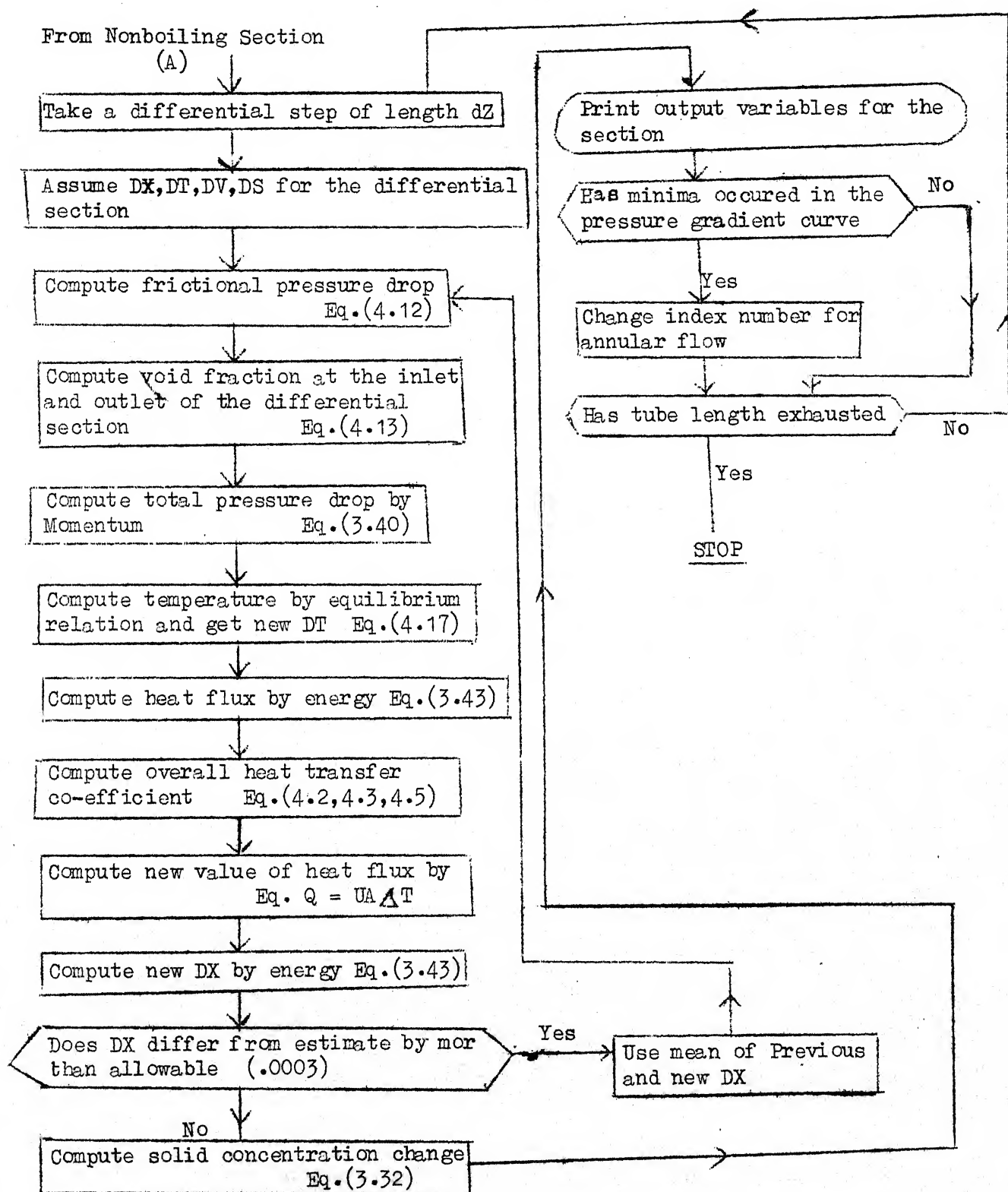
larger than a certain specified limit (.0003) then an average of these two values of quality is used and the calculations are repeated from the step computing heat flux from energy eq. 3.21 until the difference between calculated and the assumed values of DX is within specified limit. In case difference is within specified limit, solid concentration change is obtained by eq. 3.32. Now value of change in the velocity of homogeneous phase can be calculated by continuity eq. 3.25 and can be compared with the previously assumed values of DV . If the difference between these two values is larger than a certain specified limit(.0003 meter/sec) than calculations are repeated with this new value. In case difference is within specified limit, pressure gradient curve is checked for the minima. Index number which will make change from nucleate boiling to annular film flow will be changed with the occurrence of minima in the pressure gradient curve. This iterative procedure is carried out for several differential sections until whole of the length is traversed.

For annular flow region, computational procedure is same as for nucleate boiling region except Penman and Tait eq. 4.4 of annular film heat transfer co-efficient is used in place of eq. 4.2 & 4.3.

Slip Model:

Slip model implies a different velocity for liquor and vapor phases. Flow chart for the simulation of heat transfer and other phenomena based on slip model is given in Fig.5.3. This scheme of computation based on slip model is essentially the same as for homogeneous model except with a slight difference that eq. 3.40 and 3.43 are used to estimate pressure drop and temperature change respectively for the differential section.

Fig. 5.3 Flow Diagram for Boiling Section(Slip Model) of a LTV Tube



CHAPTER 6

RESULTS & DISCUSSION

The computational scheme outlined in chapter 5 is used to simulate the long tube vertical evaporator for the following liquor systems: water, bamboo black liquor, ~~lignin~~ black liquor and pinewood black liquor.

Single Effect LTV Evaporation Simulation:

A. Application of Simulation Models to Badger's Data for Water:

The accuracy of the simulation programme for single effect evaporator calculations has been compared with the experimental data of Brooks and Badges⁽⁴⁾ for the evaporation of water in a LTV unit consisting 6.1 meter long and 5.08/4.45 Cm diameter tube. Brooks and Badges's experimental data for water are summarized in Table 6.1. The variation in process variables along the tube length is determined by the simulation programme based on flow charts Fig. 5.1 and Fig.5.2 for nonboiling and boiling section (Homogeneous model) respectively. The length of the boiling section and the corresponding rates of evaporation and heat transfer, and the average overall boiling heat transfer co-efficient ($A_{o3} U = \sum U_i \Delta T_i / \sum \Delta T_i$) obtained from the simulation programme given in Table 6.2 for the experiments reported by Brooks and Badger. Brooks and Badger's results are also included in Table 6.2 for comparison purposes. The range and average percent deviation

Table 6.1-Badger's Experimental Data for the Evaporation of Water in LTF Unit

Data	Feed rate (Kg/hr)	Feed inlet (°C)	Steam temp. (°C)	Boiling length (Meter)	Evaporation rate (Kg/hr)	Heat transfer rate in the boiling section (KW)	Average overall heat transfer co-efficient in boiling section (KW/m ² °K)
1	107.2	72.1	99.0	5.65	75.0	47.80	3.26
2	108.0	50.4	79.5	5.19	43.4	27.62	3.45
3	120.2	60.0	76.0	5.25	38.8	24.40	3.57
4	112.1	62.0	96.0	5.48	73.4	47.55	3.49
5	113.6	38.0	75.0	4.45	29.3	18.10	3.13
6	121.0	38.5	82.0	4.91	59.5	37.50	3.34
7	111.4	51.0	81.0	5.12	54.6	34.66	3.25
8	112.8	60.0	81.0	5.45	56.5	35.60	3.24

L = 6.1 meter

D = 5.08/4.45 Cm.

Table 6.2 - Simulation Results for Water Evaporation Using

Homogeneous Model

Data	Boiling Section Length (meter)	Evaporation Rate (Kg/hr)	Heat Transfer rate in the boiling Section (KW)	Average overall Heat transfer Co-efficient in Boiling Section (KW/m ² °K)
Computed value	4.84	39.6	24.20	3.04
1 Experimental value	5.65	75.0	47.80	3.26
% Deviation	-14.3	-47.2	-49.5	-7.1
Computed value	4.21	73.2	20.13	3.16
2 Experimental value	5.19	43.4	27.62	3.45
% Deviation	-18.8	69.5	-26.6	-8.7
Computed value	4.28	52.8	17.62	3.07
3 Experimental value	5.25	38.8	24.40	3.57
% Deviation	- 18.6	37.8	-28.6	-10.8
Computed value	4.67	105.8	26.00	3.07
4 Experimental value	5.48	73.4	47.55	3.49
% Deviation	-15.0	43.8	-45.0	-11.9
Computed value	3.59	58.6	16.90	3.14
5 Experimental value	5.45	29.3	18.10	3.13
% Deviation	-19.6	101.0	-5.0	0.5

(Table 6.2 continued)

Computed value	4.04	91.5	22.94	3.15
6 Experimental value	4.91	59.5	37.50	3.34
% Deviation	-19.5	54.0	-39.1	-5.6
Computed value	4.25	79.5	21.44	3.17
7 Experimental value	5.12	54.6	34.66	3.25
% Deviation	-17.1	46.5	-38.3	-2.54
Computed Value	4.58	74.2	21.70	3.14
8 Experimental value	5.45	56.5	35.60	3.24
% Deviation	-16.2	32.0	-39.8	-2.8
Deviation range %	14-20	32-101	5-50	0.5-12
% Average Deviation	17.2	53.9	33.9	6.2

of the calculated values from the Brooks and Badger's experimental observations are given below (assuming homogeneous model for the boiling section):

<u>Process Variables</u>	<u>Deviation range, %</u>	<u>Average Deviation, %</u>
Boiling section length	14-20	17
Evaporation rate	32-101	54
Heat Transfer rate	5-50	34
Boiling average overall		
heat transfer co-efficient	2-12	6

The above results show that while this simulation model can predict the overall heat transfer co-efficient quite well, accuracy for predicting the boiling length can be considered moderate and the estimates of the rate of evaporation and heat transfer by the model are poor. The rather large deviations observed may be attributed to the limitations imposed by the homogenous model which assumes equal velocity for the gas and liquid phases during two phase flow.

The slip model for the boiling section was then used to simulate the process conditions of Brook & Badger's experiments based on flow chart 5.1 & 5.3. Results as obtained by using this model and Badger's experimental results are given in Table 6.3 which also include % deviation of the predicted values from the experimental values. The following observations can be made from the results in Table 6.3, concerning the accuracy of the simulation programme based on slip model for two phase

Table 6.3 Simulation results for Water Evaporation Using Slip Model:

Data	Boiling Section Length (Meter)	Evaporation Rate (Kg/hr)	Heat Transfer rate in the boiling Section(KW)	Average overall Heat transfer Co-efficient in Boiling Section (KW/m ² °K)
Computed value	5.42	68.3	43.90	2.68
1 Experimental value	5.65	75.0	47.80	3.26
% Deviation	-4.1	-8.5	-8.6	-17.5
Computed value	4.21	40.0	25.77	2.76
2 Experimental value	5.19	43.4	27.62	3.45
% Deviation	-18.8	-7.6	-6.8	-20.0
Computed value	4.28	35.0	23.00	2.78
3 Experimental value	5.25	38.8	24.40	3.57
% Deviation	-18.6	-9.4	-6.9	-21.0
Computed value	4.99	66.5	42.40	2.71
4 Experimental value	4.58	73.4	47.55	3.49
% Deviation	-7.3	-8.9	-11.1	-22.2
Computed value	3.59	33.4	21.20	2.76
5 Experimental value	5.45	29.3	18.10	3.13
% Deviation	-19.6	14.8	16.4	-11.6

(Table 6.3 continued)

	Computed value	4.04	48.8	30.94	2.72
6	Experimental value	4.91	59.5	37.50	3.34
	% Deviation	-19.5	-17.5	-18.0	-18.3
	Computed Value	4.25	44.1	28.40	2.79
7	Experimental value	5.12	54.6	34.66	3.25
	% Deviation	-17.1	-18.9	-16.9	-14.1
	Computed value	4.58	46.0	28.95	2.78
8	Experimental value	5.45	56.5	35.60	3.24
	% Deviation	-16.2	-18.4	-20.2	-14.2
	Deviation range %	4- 20	8- 19	6- 2'	10- 22
	% Average Deviation	15- 2	13.2	13.1	17.2

flow:

<u>Process Variables</u>	<u>Deviation range, %</u>	<u>Average Deviation, %</u>
Boiling section length	4 - 20	15.2
Evaporation rate	8 - 19	13.2
Heat Transfer rate	6 - 20	13.1
Boiling average overall		
heat transfer co-efficient	10 - 22	17.2

The above results show that the simulation programme of this study assuming slip model can predict Brooks and Badger's experimental conditions with an overall accuracy of about 15%. Hence the simulation programme based on flow charts 5.1 and 5.3 for the nonboiling and boiling section (slip model for two phase flow) respectively ~~can be used~~ for the subsequent calculations of this investigation for evaporation of bamboo, baggase and pinewood black liquor from alkaline pulping processes.

B. Comparison with Gudmundson's Experimental curves:

The simulation programme is also used for comparing the predicted values with the experimental results of Gudmundson^(2,45). The evaporator used in the latter work consisted of four 10 m long stainless steel tubes with diameters of 51/46 mm. Gudmundson's results for water at a feed rate of 18 Kg/Sm² are used in this comparison.

Figures 6.1, 6.2 and 6.3 give the predicted temperature profile and overall heat transfer Co-efficient along the tubelength for the various inlet feed temperatures of 76°C, 81.5°C and 85°C; and corresponding steam temperatures of 91.5°C, 95°C and 86.5°C. The simulation calculations are represented by the continuous curves in the above figures assuming slip model for the two phase flow region. The predicted temperature

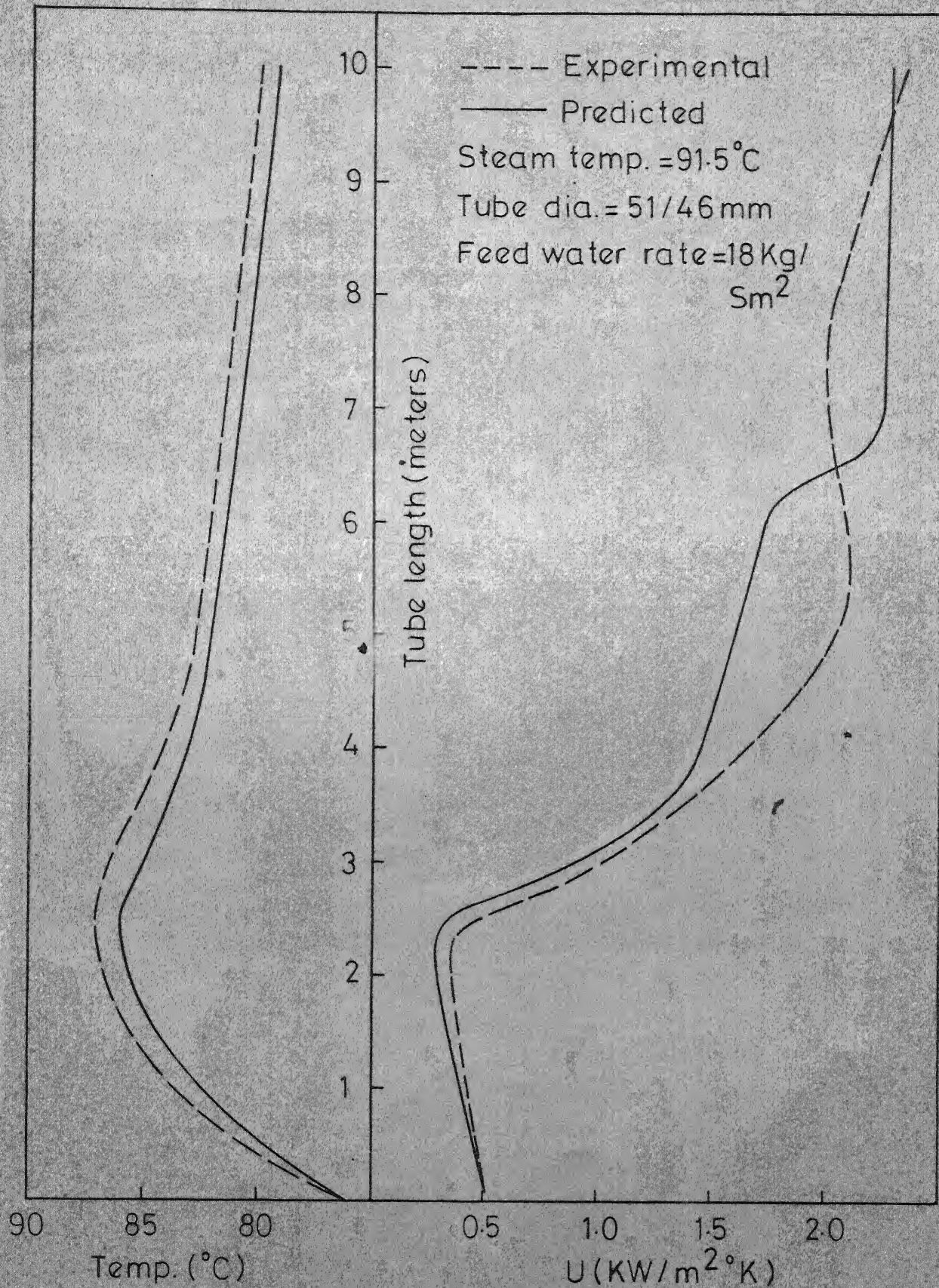


Fig.6.1 - Comparison of predicted values with experimental data for water.

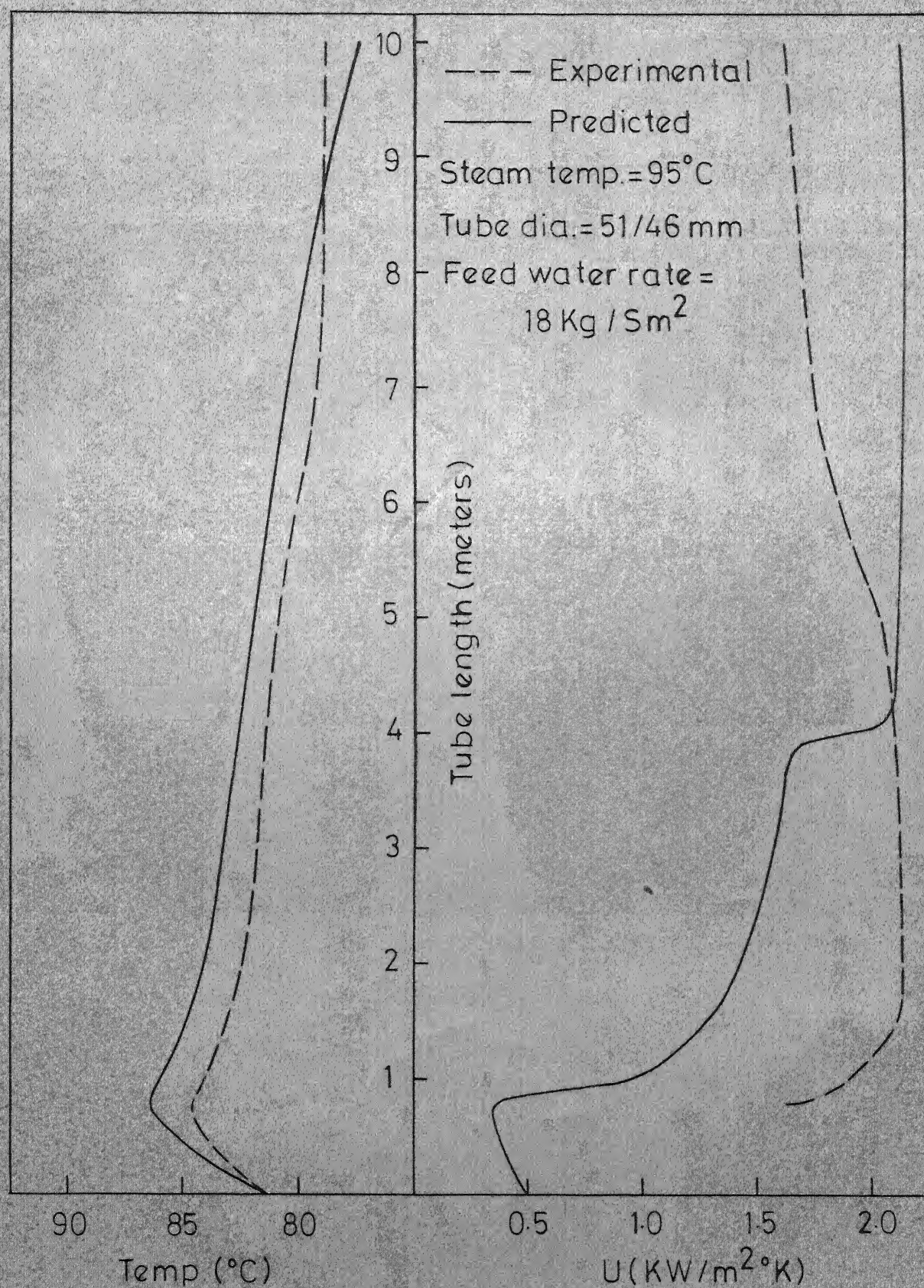


Fig.6.2 - Comparison of predicted values with Gudmundson's experimental data for water.

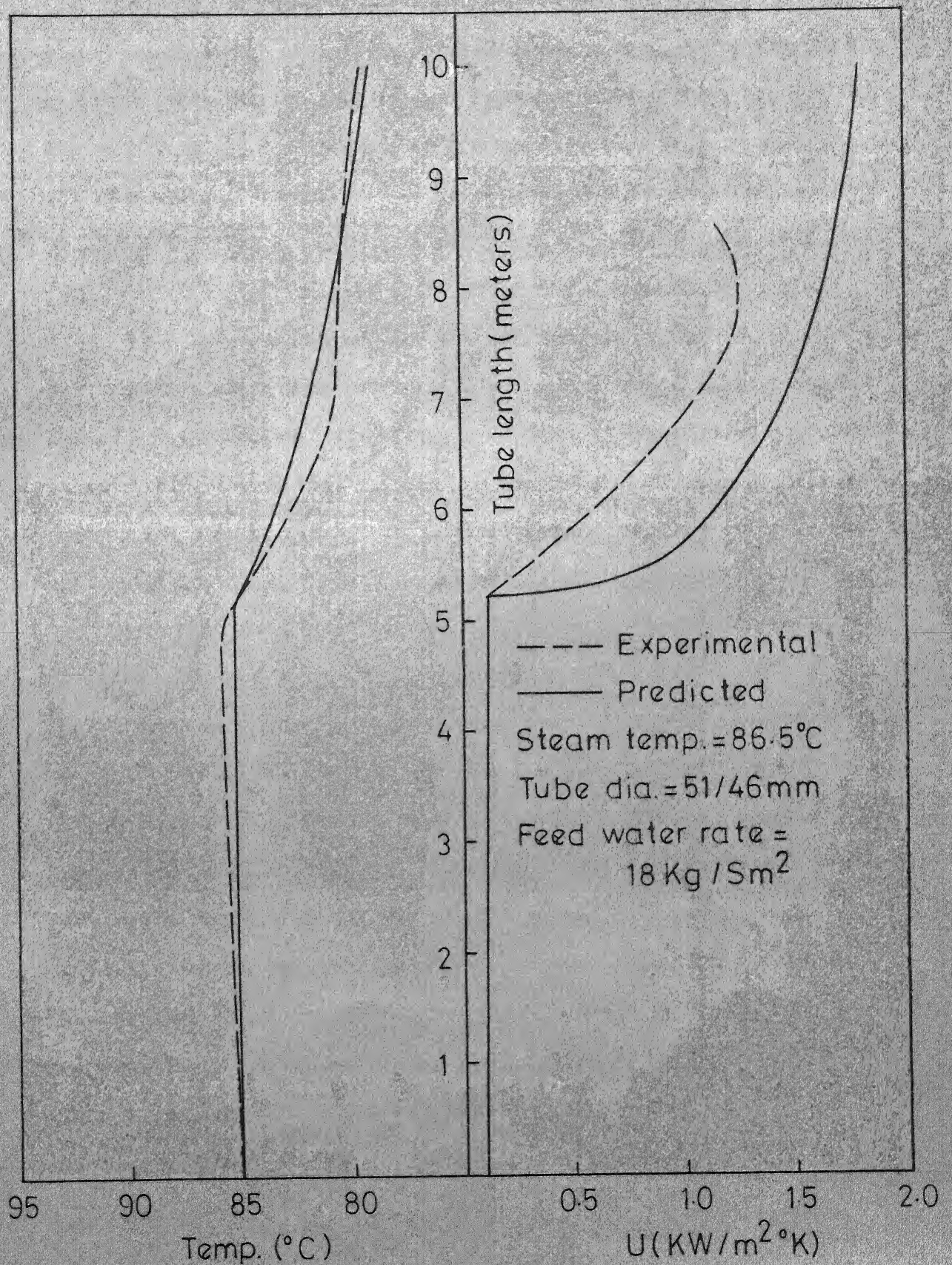


Fig.6.3 - Comparison of predicted values with Gudmundson's experimental data for water.

profiles agree with the experimental data quite well; the agreement is within 1°C for the two sets of data represented in figures 6.1 and 6.3 and is within 2°C for the data in Figures 6.2. Experimental curves for the over all heat transfer co-efficient deviates somewhat from the predicted profiles based on this study. Experimental results in general show an average deviation of about 25% from the predicted results.

The above comparisons for the heat transfer of LTV evaporator show that the simulation model of this investigation can be used to predict reliable values of heat transfer rates. Hence the simulation programme based on flow charts Fig. 5.1 and 5.3 for the nonboiling and boiling section (slip model for two phase flow) respectively are used for the subsequent calculations of this investigation for evaporation of bamboo, bagasse and pinewood black liquors.

C. Evaporation of Pinewood Black Liquor:

Weak black liquor from the Kraft pulping of pine is assumed to contain 12.0% dissolved solids with a flow rate of 100 tons/hr at a temperature of 50°C . The LTV evaporator unit is arbitrarily assumed to consist of 300 tubes, 5.08 / 4.45 cm dia. and 7.6 meters long. Steam chest pressure is taken to be 306 KN/sq.m. Pressure of the liquor feed is assumed to be 104 KN/sq m. A print out of the simulation programme for the evaporation of pinewood black liquor is given in Appendix D-1 for the representation of homogenous model describing two phase flow pattern . Physical property correlations reported by Cudmundson⁽³⁵⁾ are used in this analysis.

The simulation programme gives the output variables in the following order- vapor quality, void fraction, pressure gradient, pressure, temperature, heat flux, overall heat transfer co-efficient and solid concentration. The variations of these variables along the length of the evaporator tube are shown in Fig. 6.4 - 6.11¹ respectively. It can be observed that of the total length of the evaporator tube (7.6 meter), about 2 meter is used in raising the feed liquor to its boiling point in the non-boiling region.

Fig. 6.4 shows the variation in vapor quality (mass fraction of vapor in the total vapor -liquid mixture flow) along the tube length. It can be seen that the mass fraction of vapor increases from zero at the start of the boiling section to 0.35 and 0.22 towards the end of tube assuming homogeneous and slip models respectively. The difference observed between the two models is caused by the contribution of the kinetic energy terms (eq.3.43). In the homogeneous model kinetic energy is based on the use of homogeneous phase velocity whereas the slip model takes into account the contribution to kinetic energy as terms representing vapor and liquid velocities during two phase flow.

Fig. 6.5 gives the changes in void fraction along the tube and it can be seen that the start of nucleate boiling section is accompanied by a steep increase in void fraction caused by rapid rate of nucleation and bubble coalescence. The homogeneous and slip models give essentially the same value for void fraction. Void fraction at the end

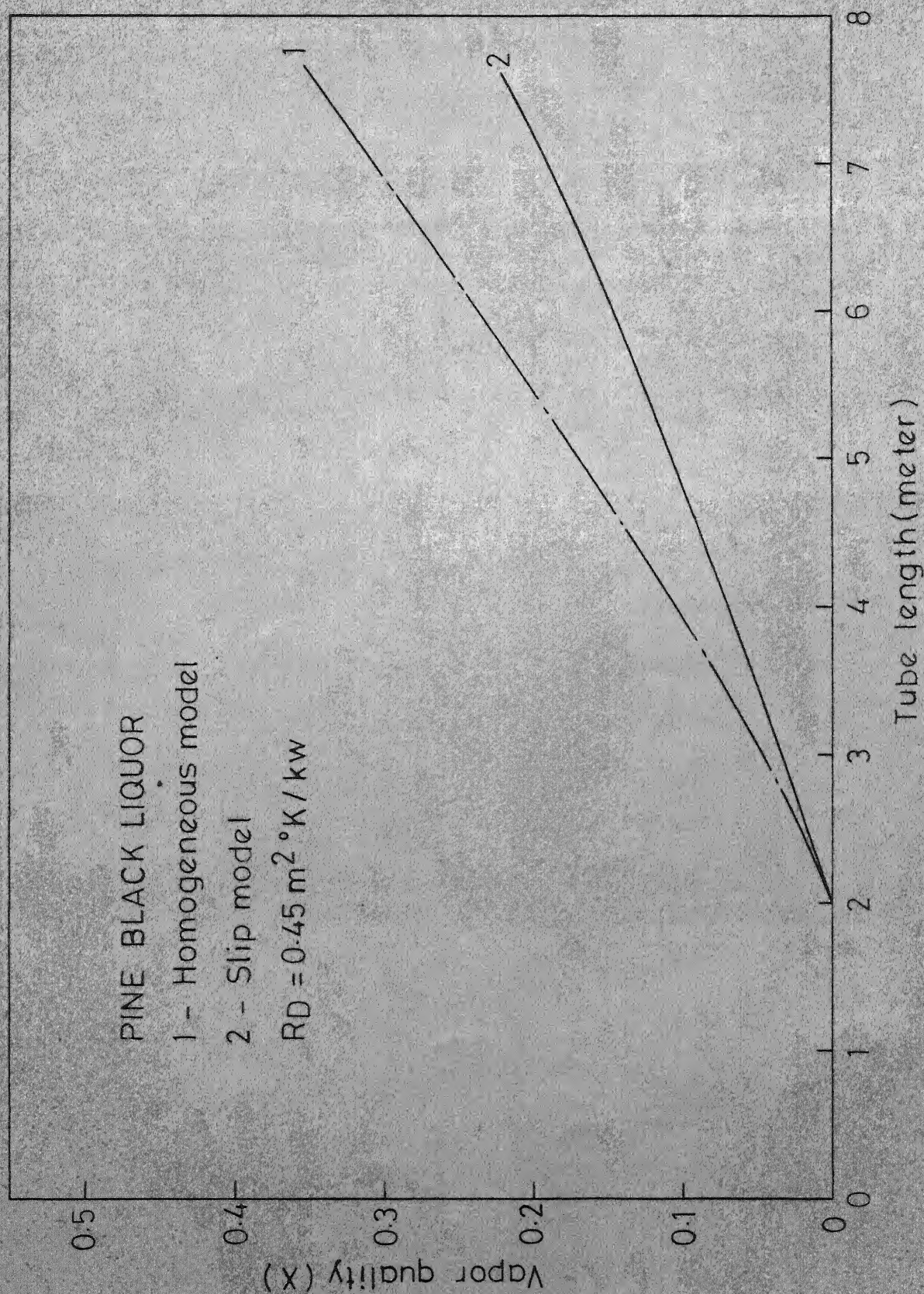


Fig.6.4- Variation of vapor quality along tube length.

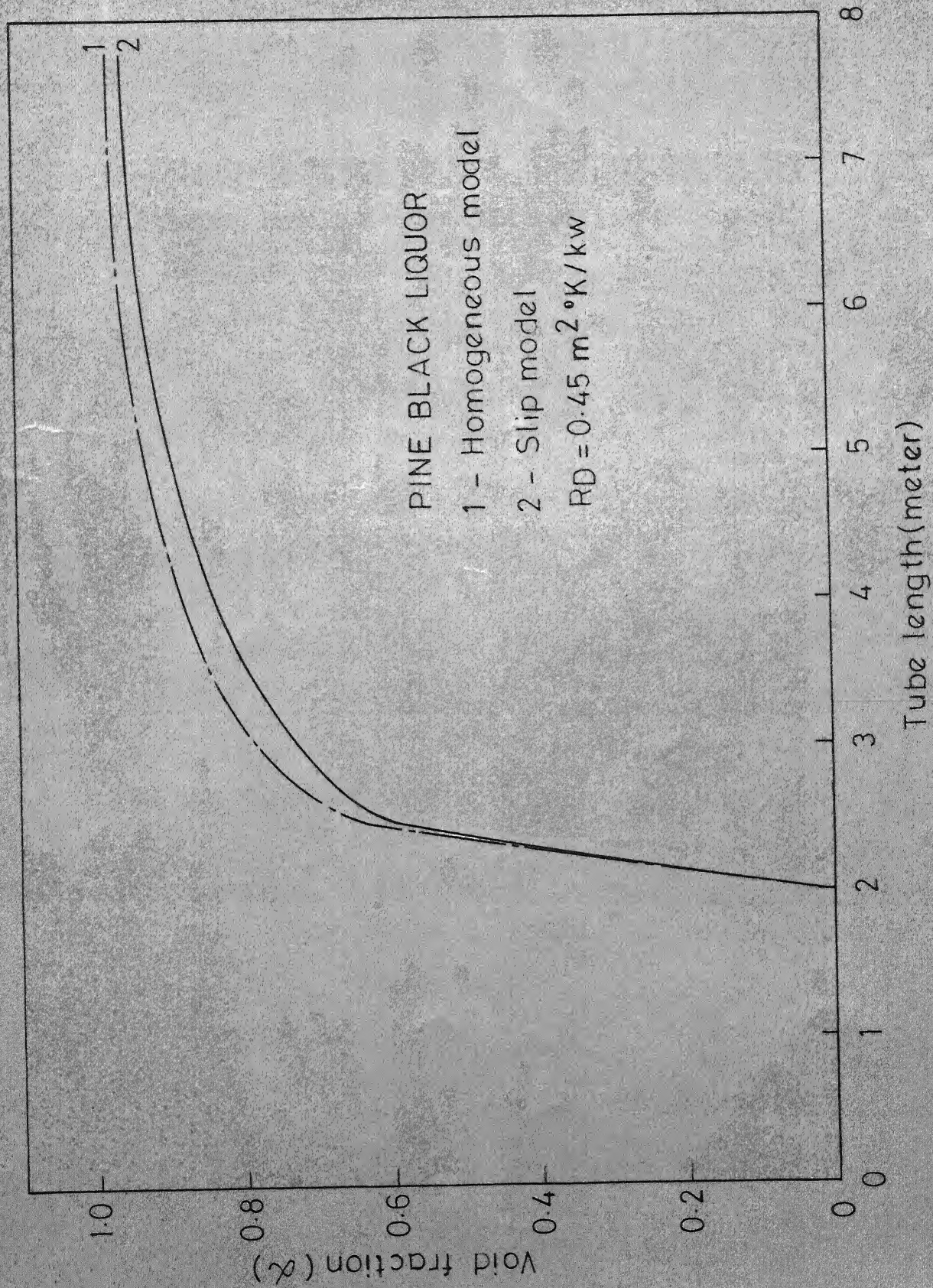


Fig.6.5- Variation of void fraction along tube length.

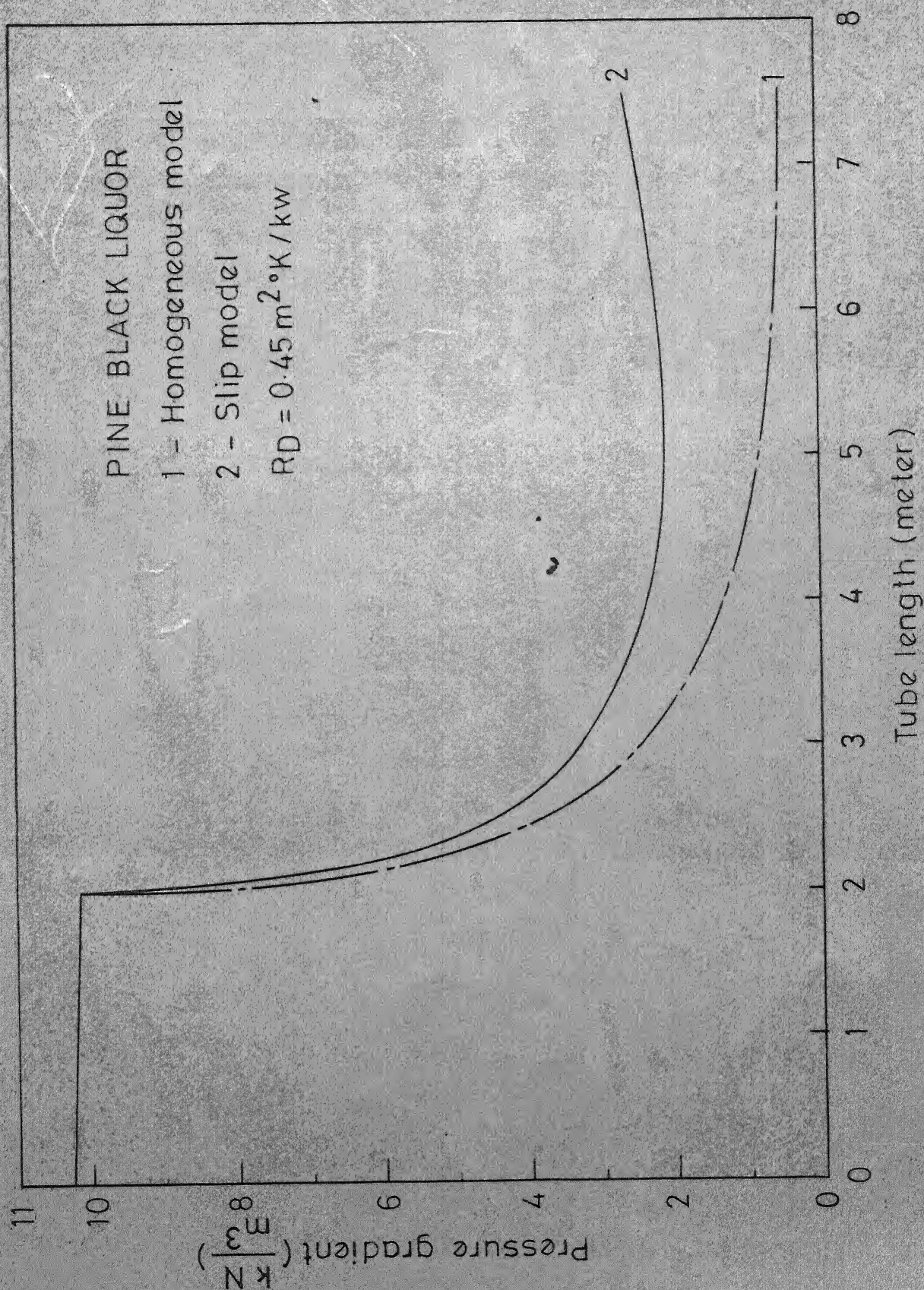


Fig 6.6-Variation of pressure gradient along tube length.

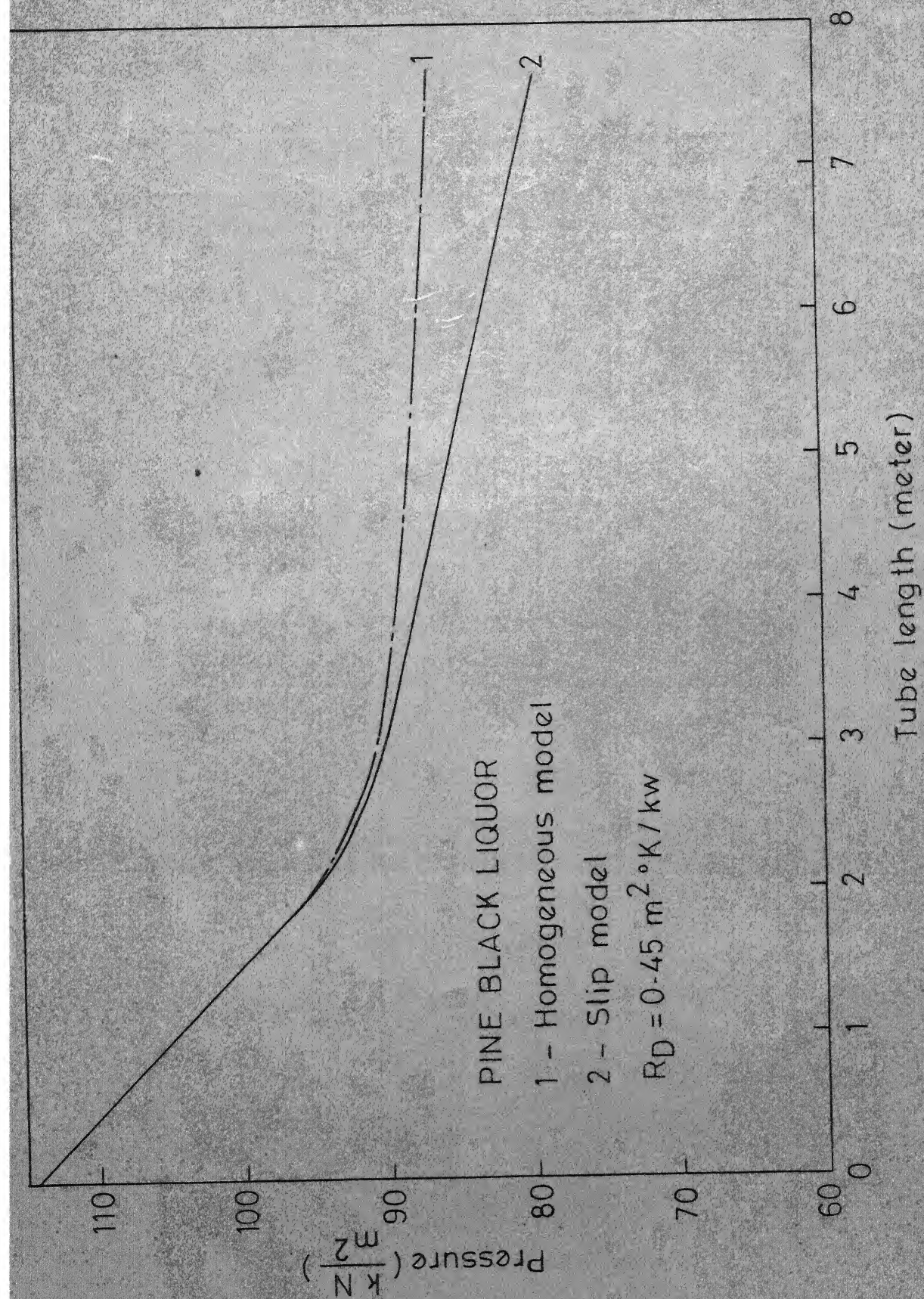


Fig.6.7- Variation of pressure along tube length.

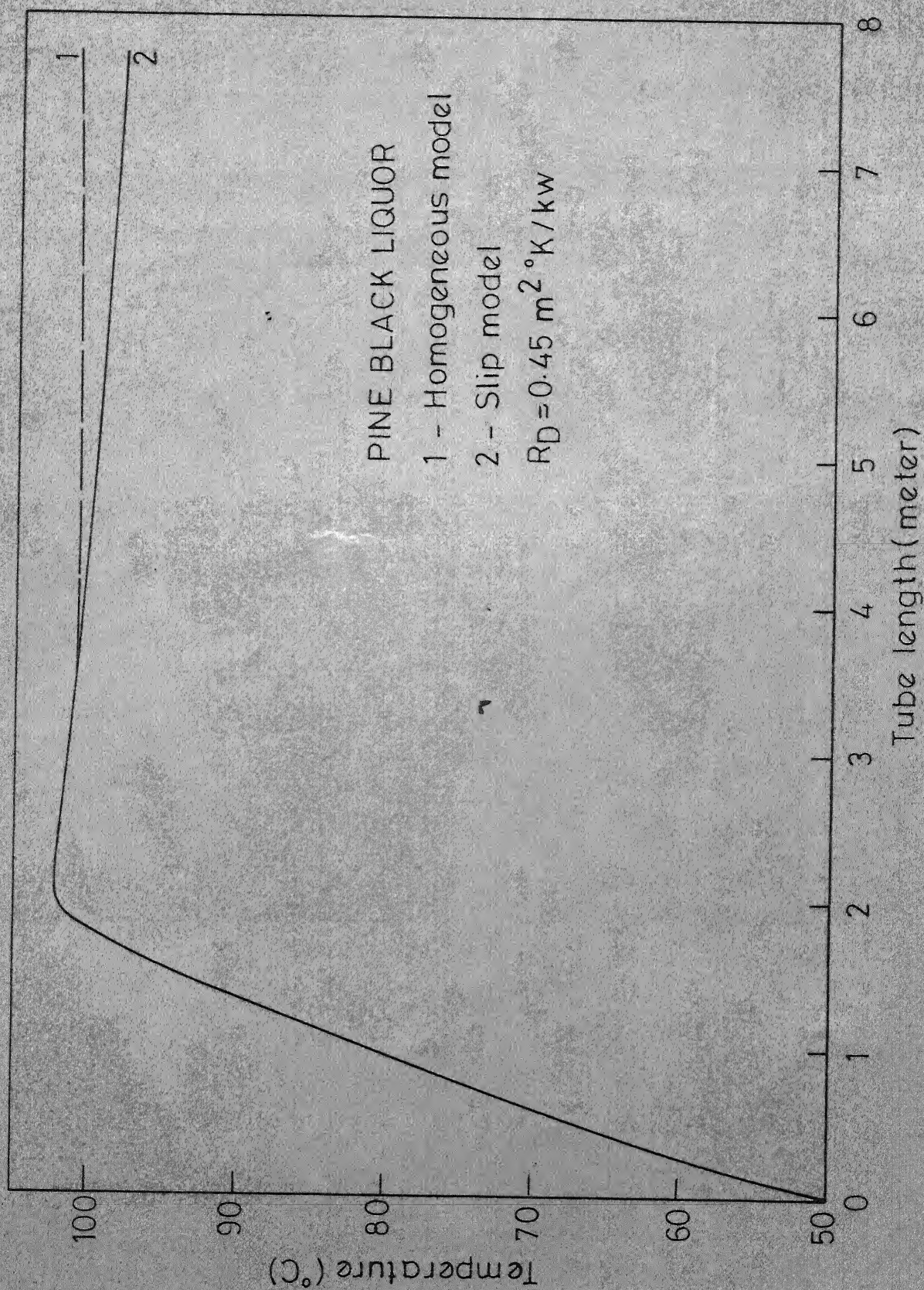


Fig. 6.8 - Variation of liquor temperature along tube length.

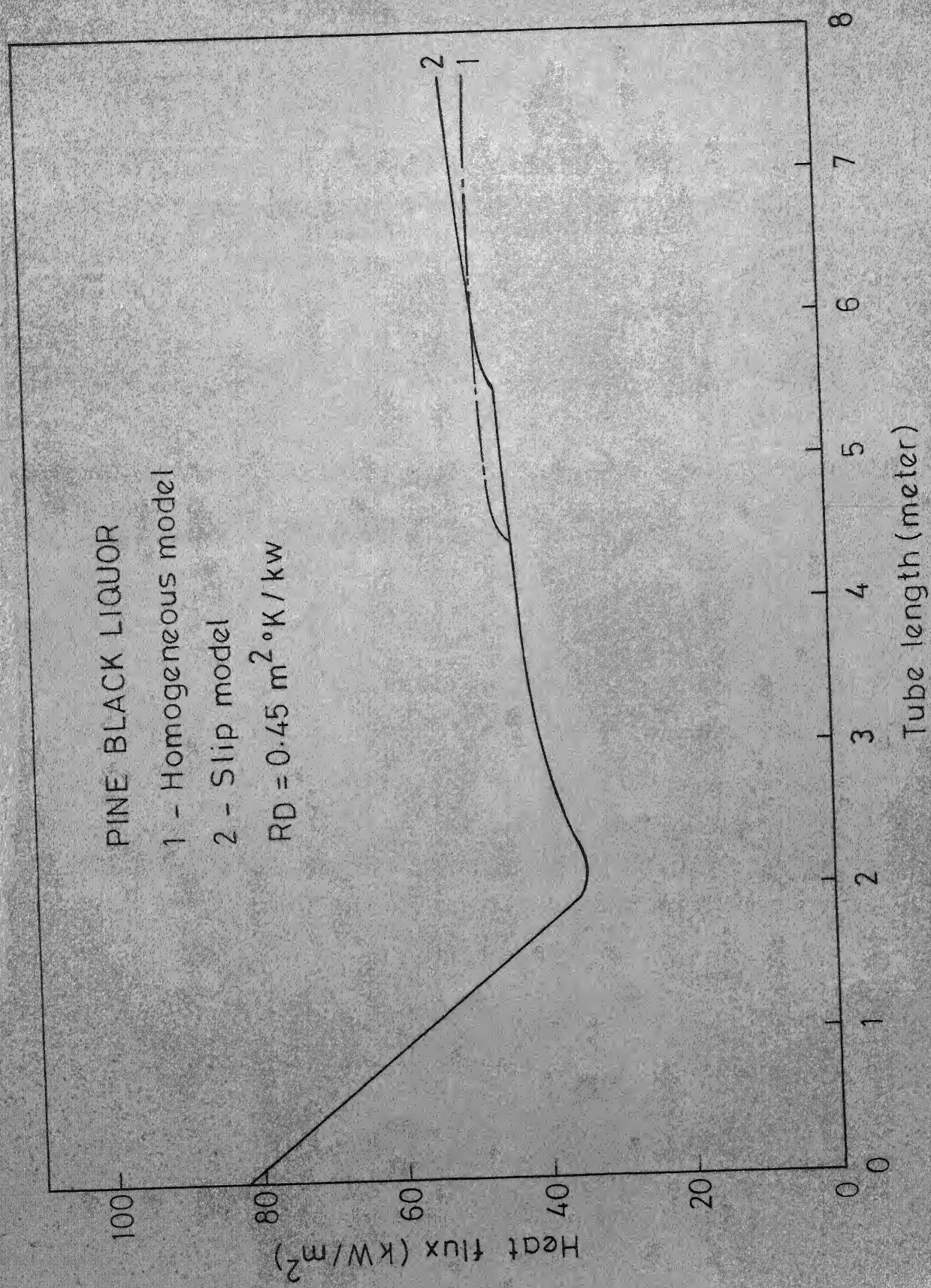


Fig.6.9- Variation of heat flux along tube length.

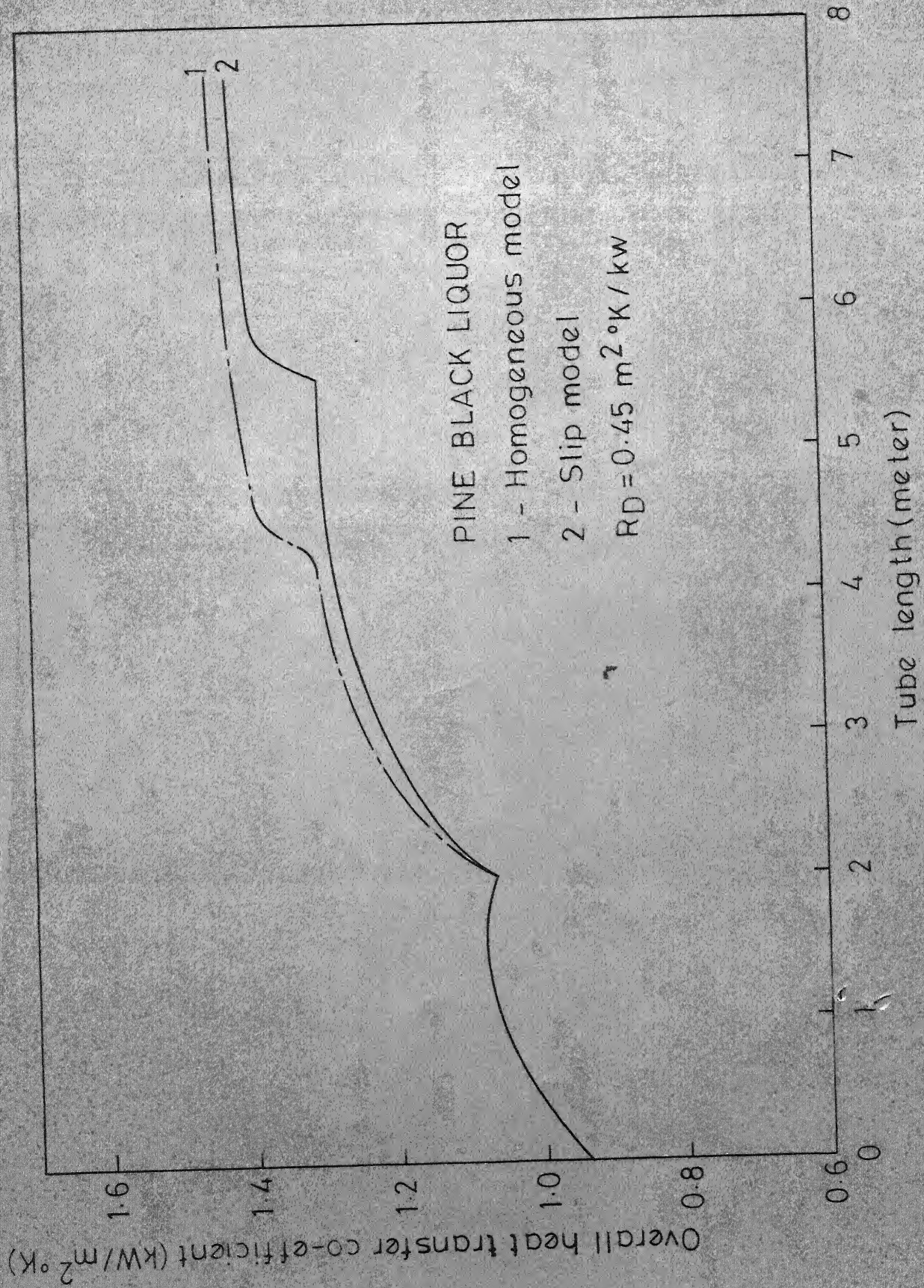


Fig 6.10-Variation of overall heat transfer co-efficient along tube length.

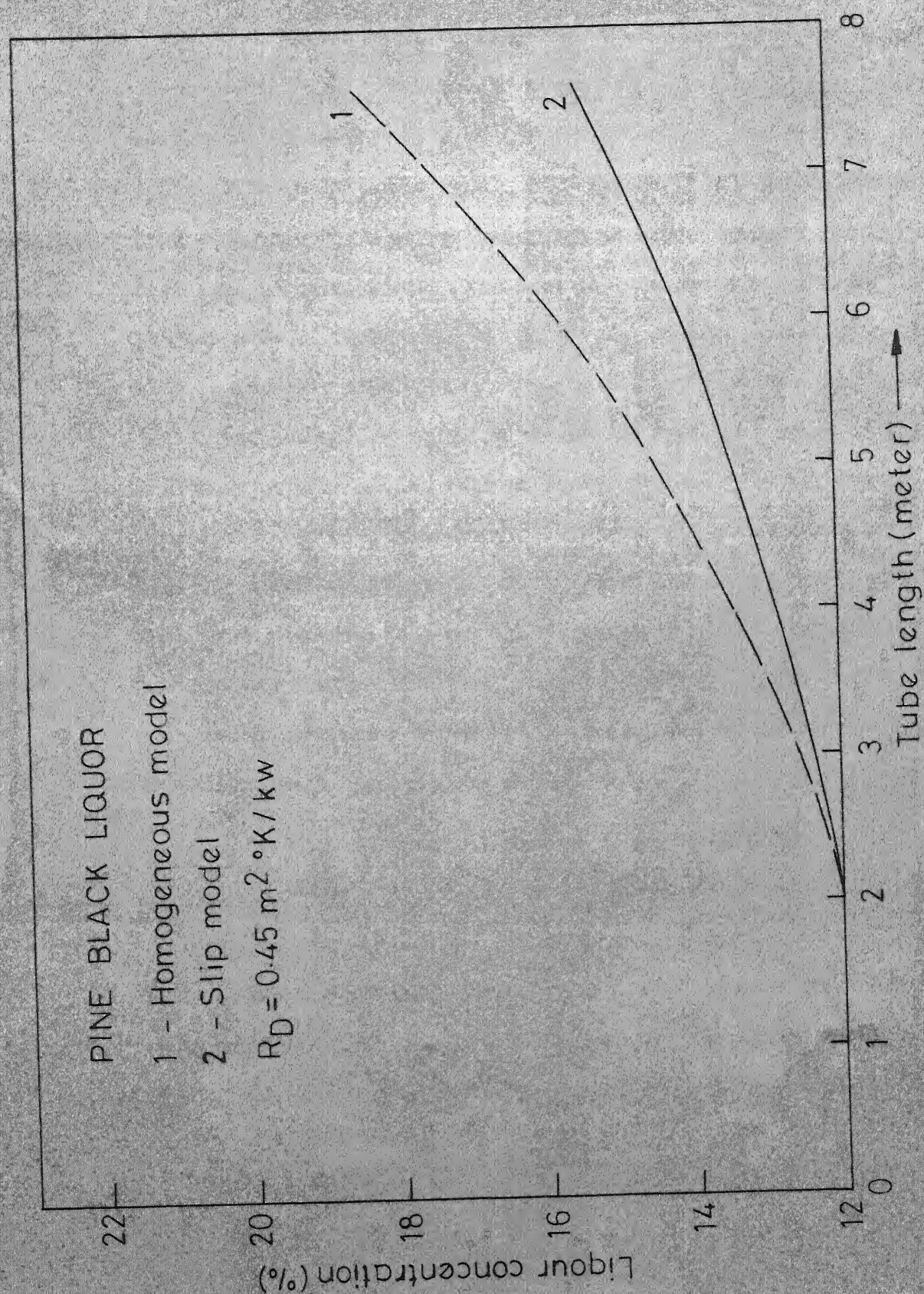


Fig.6.11 - Variation of liquor concentration along tube length.

of the evaporator tube is in the range of 0.95 - 0.97.

Change in the pressure gradient along the tube length is given in Fig. 6.6. Pressure gradient decreases slightly in the nonboiling section and subsequently rapidly in the boiling section. The observed decrease in the pressure gradient is due to static and frictional pressure drop effects. The lower value of the pressure gradient for the homogeneous model can be attributed to the fact that in homogeneous model, changes in the momentum of fluid is based on the use of homogeneous phase velocity whereas the slip model takes into account the vapor and liquid velocities during two phase flow to calculate change in the momentum of fluid. In the upper portion of the tube pressure gradient becomes substantially constant for the homogeneous model whereas it passes through a minima in the slip model which can be attributed to increasing frictional pressure drop in the annular film region. This minima in the pressure gradient corresponds to transition from nucleate boiling to annular film region. The pressure gradient curve for the slip model is similar to experimental results reported by Anderson et.al⁽¹⁵⁾.

Fig. 6.7 shows a gradual decrease in pressure and is caused by static and frictional pressure drop contributions. The steady decline in pressure in the nonboiling section is due to static pressure drop. In the boiling section pressure decreases at a lower rate due to combined effects of hydrostatic head and frictional losses. The curves in Fig. 6.7 are similar to experimental curves reported by Anderson et.al.⁽¹⁵⁾.

Fig. 6.8 shows variation of liquor temperature in the tube length. It can be seen that less than one third of the evaporator tube length is used in raising the feed from 50°C to 102.5°C (saturation temperature corresponding to the pressure in the tube) in the nonboiling section. In the boiling section extending over more than two thirds of the evaporator length. The homogeneous model and slip model for the two phase flow predict liquor temperature within 3°C of each other.

Heat flux along the tube length is given in Fig. 6.9 and the values predicted by the homogeneous and slip model are nearly equal. The decrease in the value of heat flux for the nonboiling section is caused by the decrease in temperature driving force as the liquor is being heated to its boiling point. The increasing trend in heat flux over the boiling section is due to the increase in overall heat transfer co-efficient for the nucleate boiling and annular film regions. The transition from nucleate boiling to annular film region occurs at 4.3 meter and 5.5 meters from the results of heat flux for the homogeneous and slip models assumptions respectively.

Fig. 6.10 gives the variation of overall heat transfer co-efficient along the tube length and shows clearly the three dominant heat transfer mechanisms in a LTV evaporator tube. Out of a total of 7.6 meters long tube the nonboiling, nucleate boiling and annular film regions extend over 2.0, 2.3 and 3.3 meters respectively for the homogeneous model and the corresponding lengths for the slip model are 2.0, 3.5, 2.1 meters. The transition from nonboiling to nucleate boiling and nucleate boiling

to annular film regions are accompanied by a pronounced increase in the value of overall heat transfer co-efficient. In general the homogeneous model predicts higher value for overall heat transfer co-efficient compared to the value based on the slip model. It can be attributed to the higher value of vapor quality for the homogeneous model which results in higher vapor velocity and this leads to higher heat transfer co-efficient than the slip model (eq.4.2,4.3& 4.4).

Fig. 6.11 gives the change in liquor concentration along the tube length and predicts an increase from a initial concentration of 12% to 18.4% and 15.4% at the tube exit end for the homogeneous and slip models respectively. The higher concentrations observed for the homogeneous model is due to the higher value of overall heat transfer co-efficient obtained in Fig.6.10 compared to slip model.

D. Evaporation of Bamboo Black Liquor:

The simulation programme following the flow charts in Fig. 5.1 and 5.3 based on the slip model for two phase flow is used for the evaporation of bamboo black liquor in a single effect LTV evaporator. Physical property correlations reported by Koorse⁽³⁸⁾ are used in these calculations. Evaporator dimensions and process conditions are chosen to be the same as that used for pinewood black liquor in section C earlier. Two cases of evaporation are assumed for the calculations namely (1) clean tube (2) tube with a dirt resistance of $0.45 \text{ m}^2 \text{ }^\circ\text{K/KW}$. The dependence of the variable on tube length is given in Fig.6.12-6.19 for vapor quality, void fraction, pressure gradient, pressure, temperature, heat flux overall heat transfer co-efficient and solid concentration.

The comparison of the curves 6.12-6.19 with those in Fig.6.4-6.11 shows that the nature of the curves are very similar for the evaporation of Bamboo and pine black liquors. It can be seen from Fig.6.12-6.19 that the contribution of the dirt scale resistance is an important factor and significantly reduces the value of solid concentration obtained for the tubes with clean heat transfer surfaces. Dirt scale resistance increases the tube length corresponding to nonboiling region while decreasing the tube length available for boiling region. For the process conditions assumed for the evaporation of black liquor, the nonboiling section prevailing over 1 meter length of clean tube increases to 2 meter when dirt scale resistance is assumed to $0.45 \text{ m}^2 \text{ }^\circ\text{K/KW}$.

Fig. 6.12 gives variation of vapor quality with tube length and it can be seen that the quality increases from zero at the start of the boiling section to 0.55 and 0.23 towards the end of the tube for clean and fouled surfaces respectively. The higher vapor quality obtainable for the clean surfaces suggest that the process liquor can be evaporated to a higher solid concentration for the given length of the tube. Fouled surface have the effect of decreasing the capacity obtainable using cleaner evaporator tubes.

The graph obtained for the void fraction in Fig.6.13 for bamboo black liquor is essentially identical to the curves obtained for pine black liquor in Fig. 6.5 assuming equal dirt resistance in the tube.

Fig. 6.14 gives the variation of pressure gradient along the evaporator tube. For the last part of the tube length value of pressure

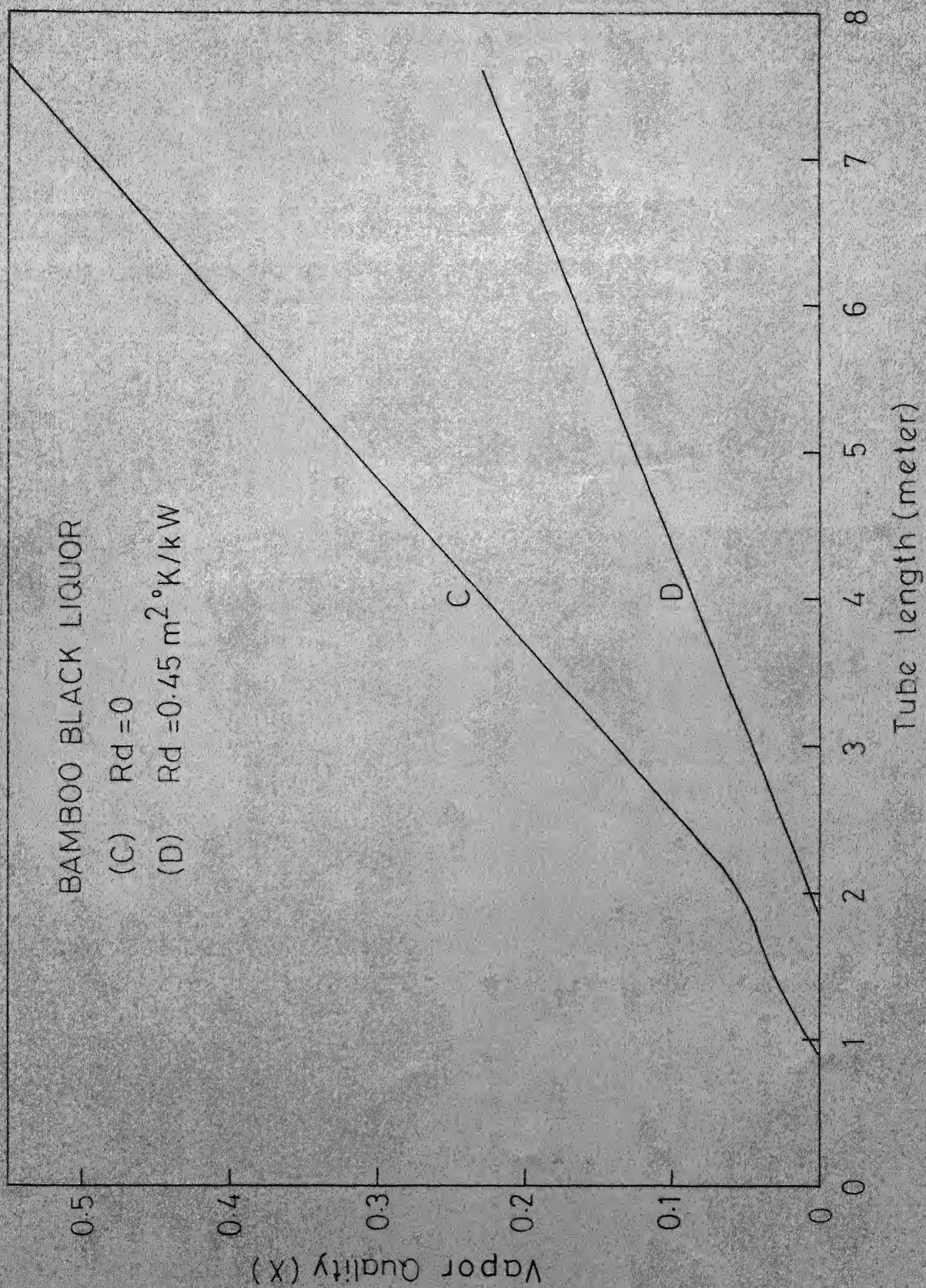


Fig.612-Variation of vapor quality along tube length.

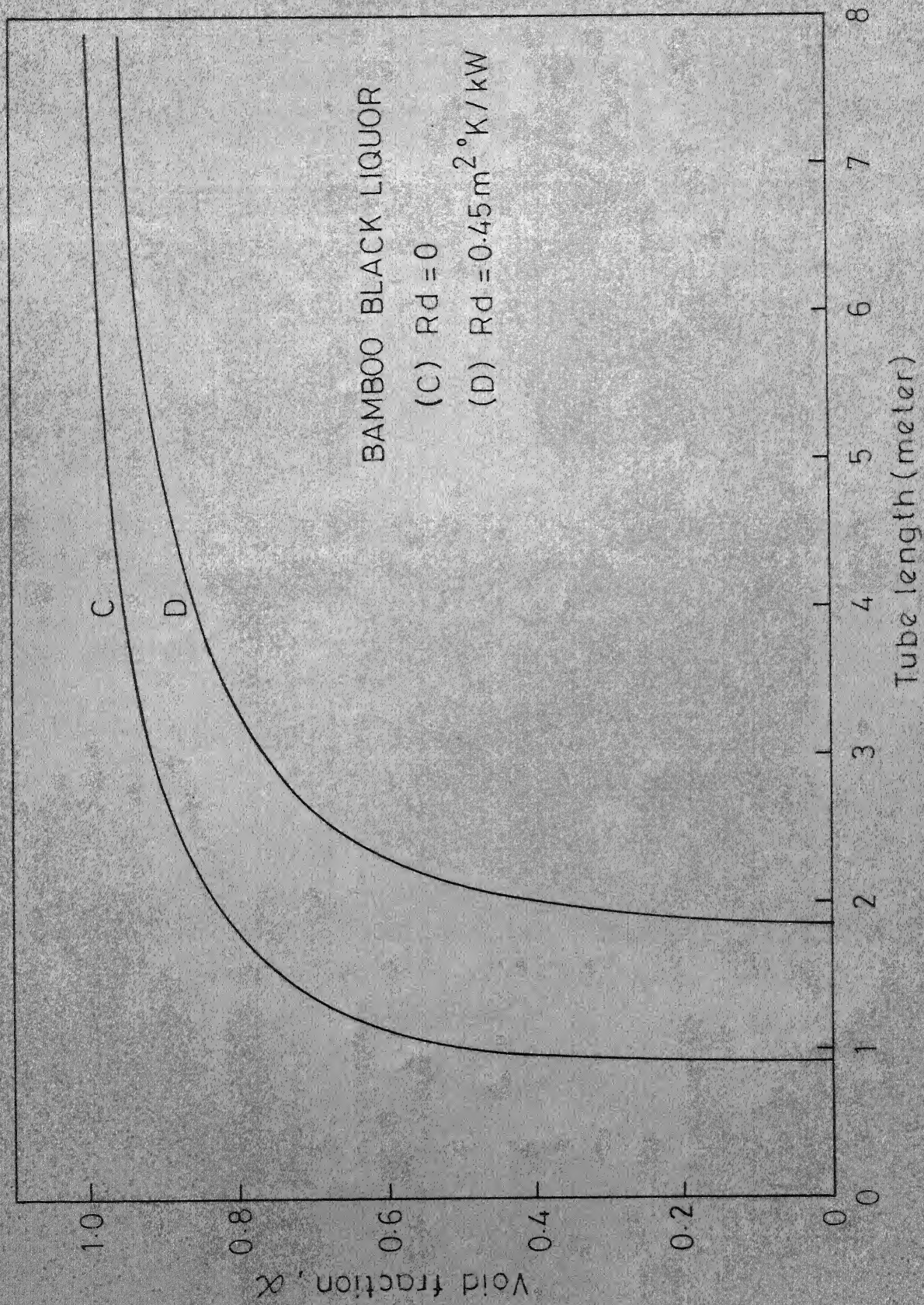


Fig 613-Void fraction in fluid flow along tube length.

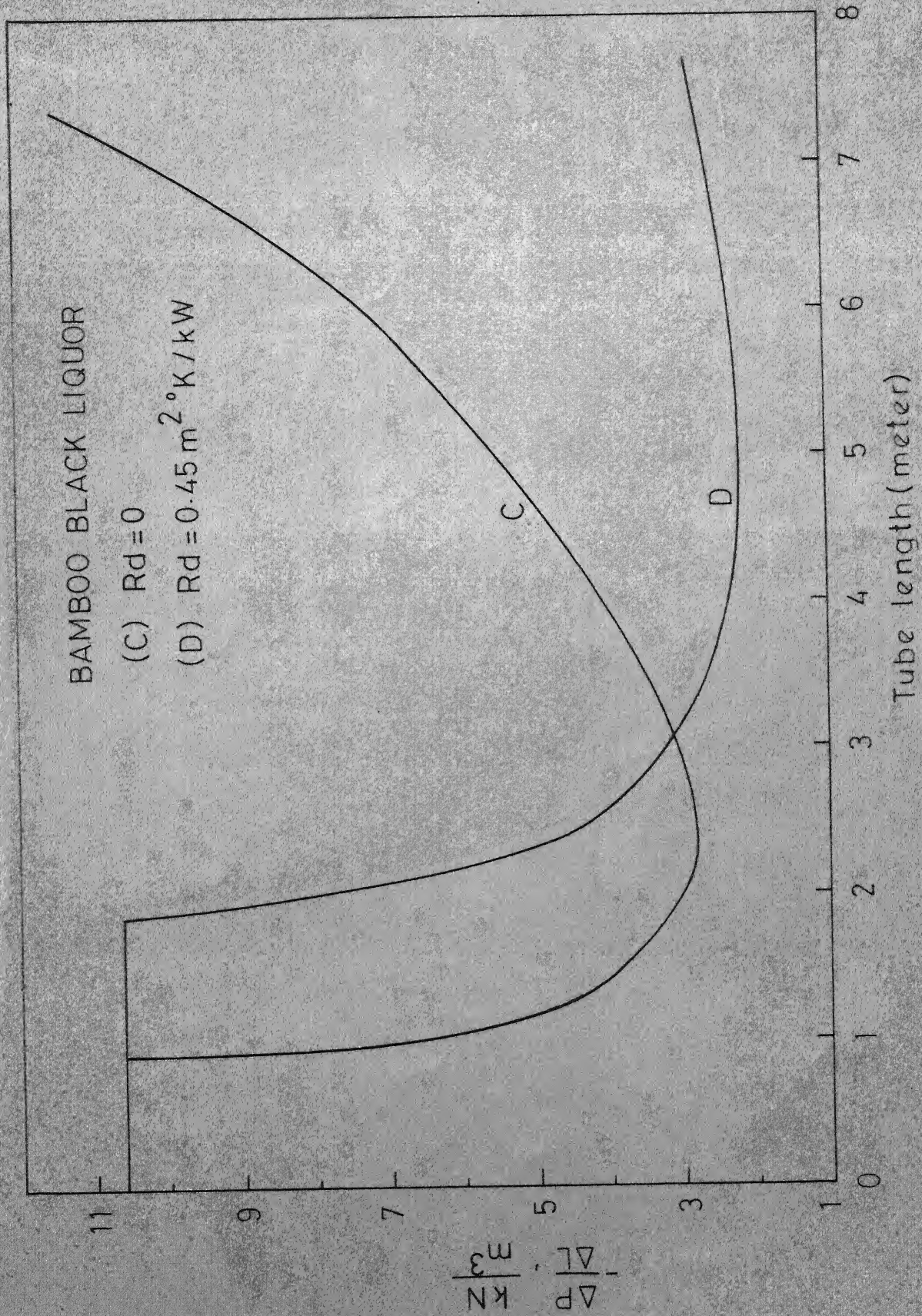


Fig.6.14- Variation of pressure gradient along tube length.

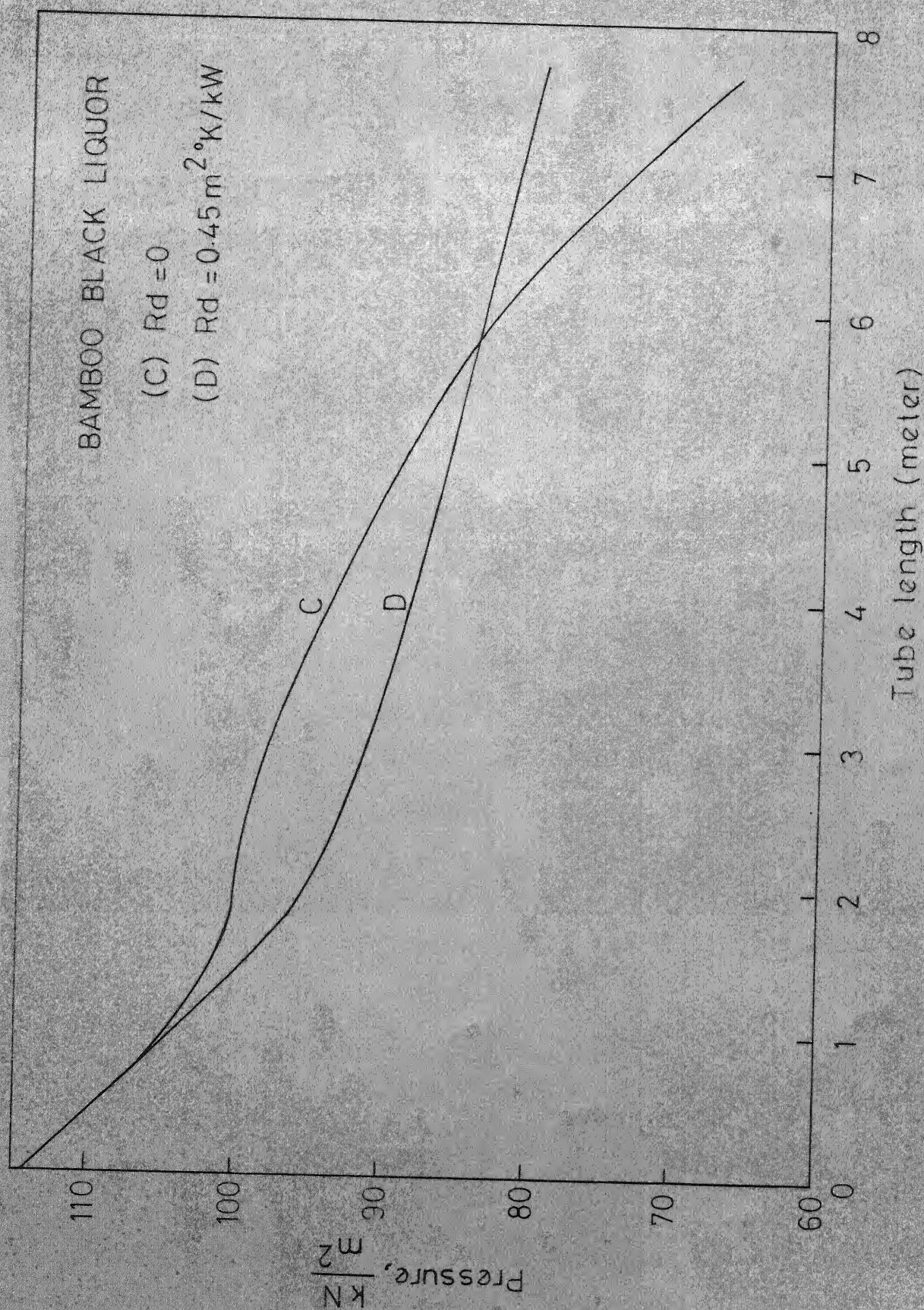


Fig.6.15- Variation of liquor pressure along tube length.

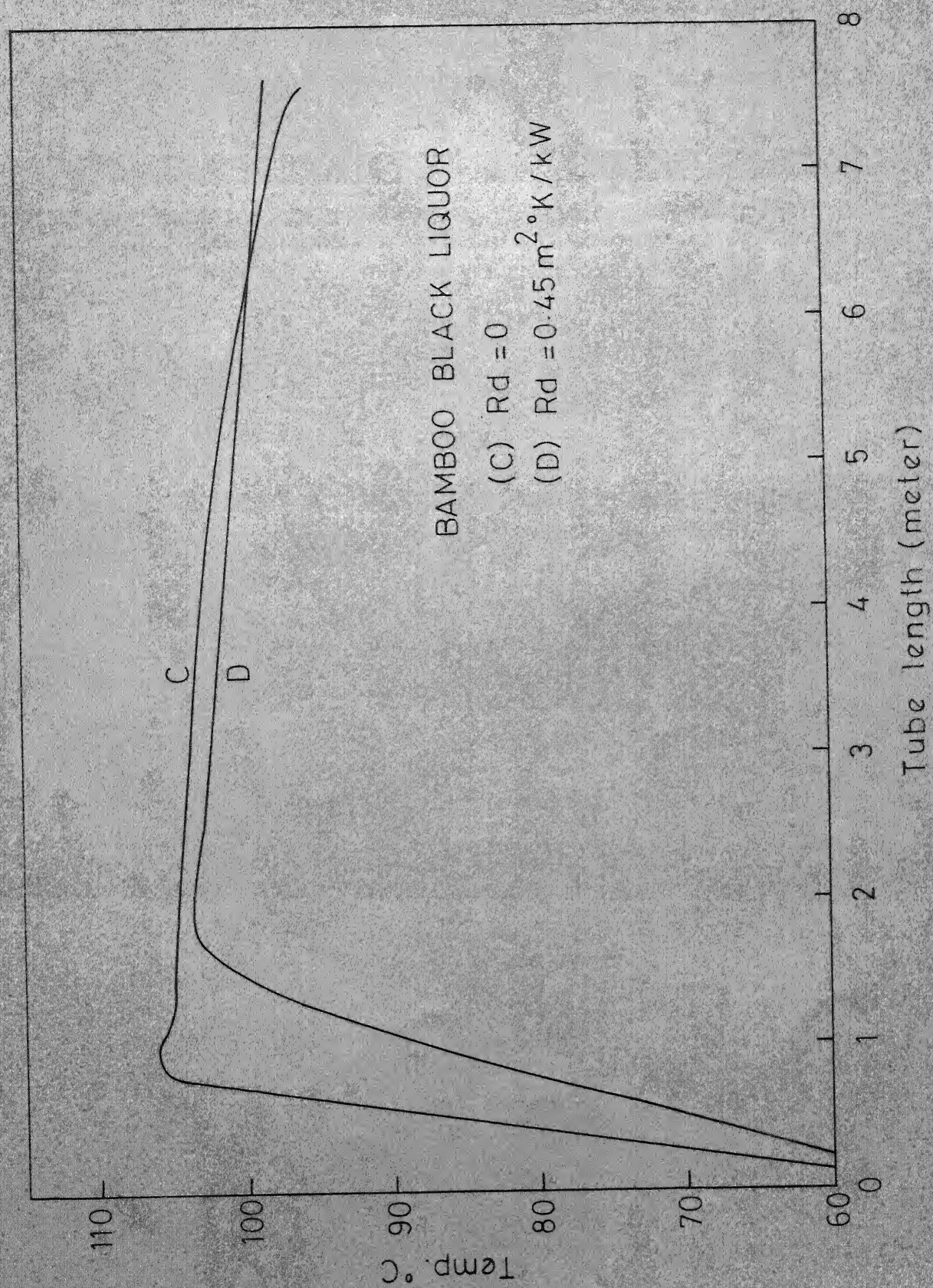


Fig.6.16- Variation of liquor temperature along tube length.

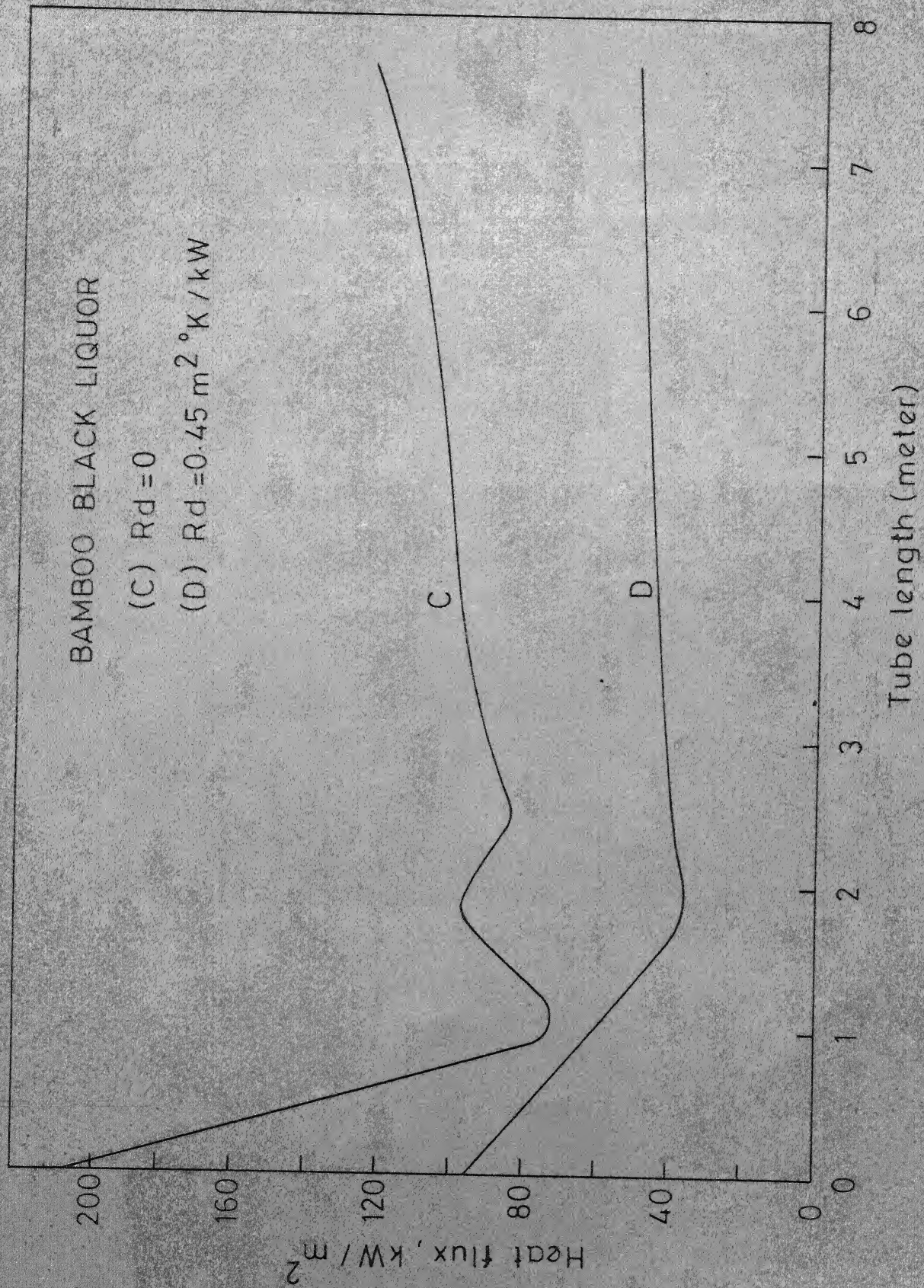


Fig.6.17- Variation of heat flux along tube length.

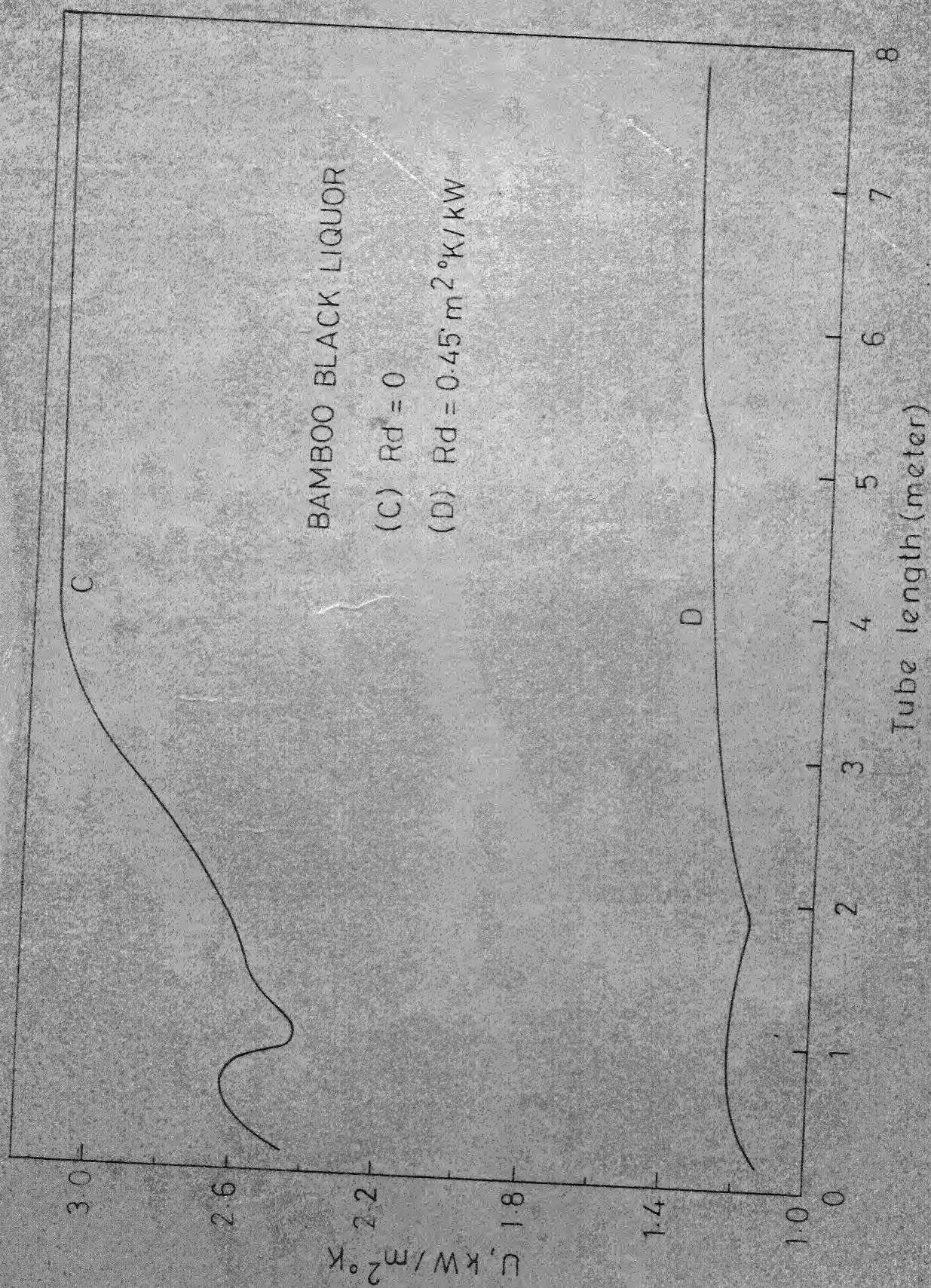


Fig.6.18- Variation of overall heat transfer coefficient along tube length

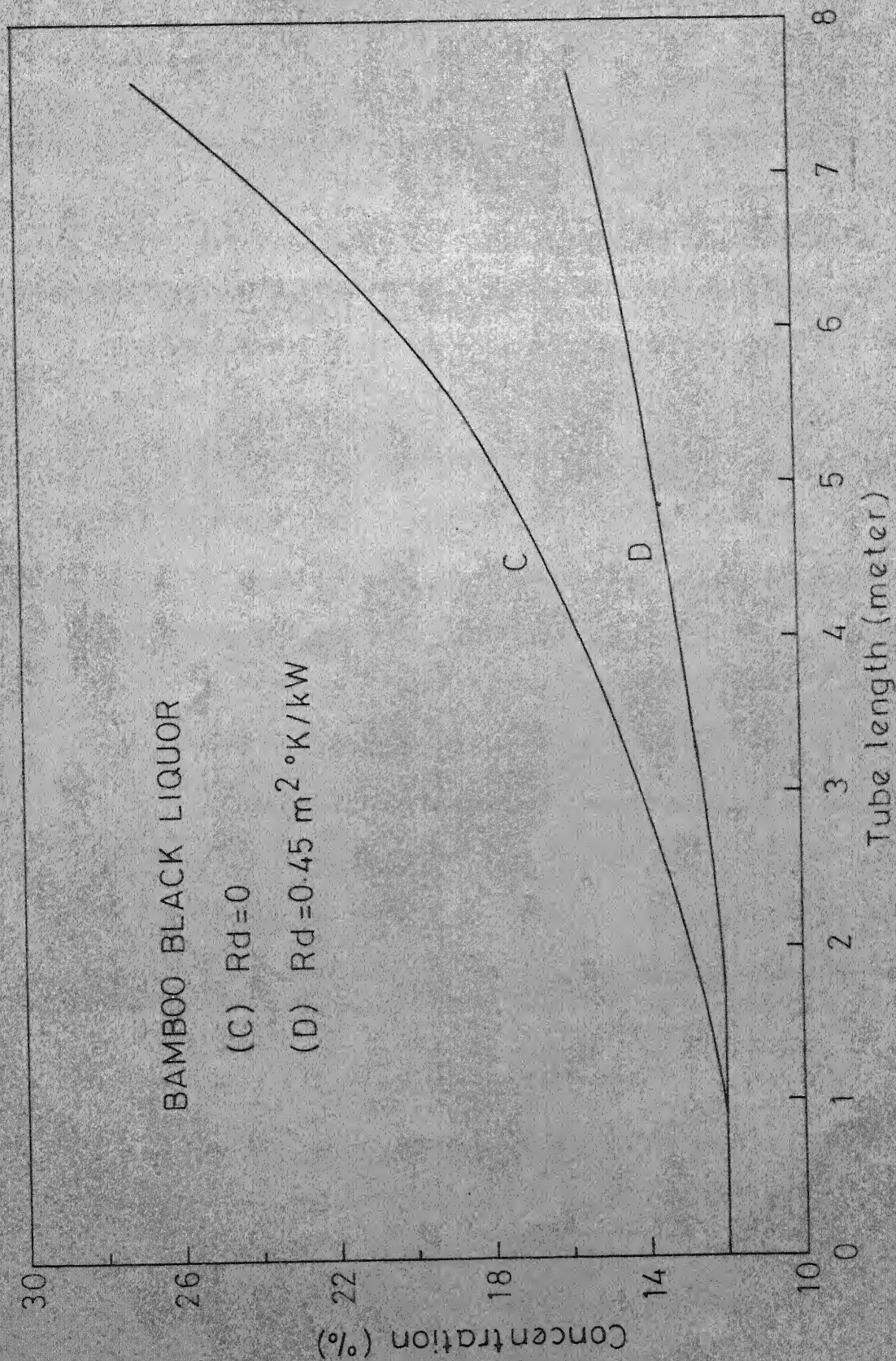


Fig 6.19- Variation of liquor concentration along tube length.

gradient is higher for the clean tube in comparison to fouled tube. The curves go through a minima at 2.2 and 5.2 meters for the clean and fouled tubes respectively and these lengths represent the transition from nucleate boiling to annular film regions. The pressure gradient in the annular film region is up to 5 times the value for the fouled tubes having the dirt resistance of $0.45 \text{ m}^2 \text{ }^\circ\text{K/KW}$. Such a condition can cause regions of low absolute pressure in the annular film region which may lead to plugging of tubes.

Fig. 6.15 describes the longitudinal variation of pressure along the tube length based on pressure gradient given in Fig. 6.14.

Fig. 6.16 gives the profile of liquor temperature and shows that the maximum value of $106.0 \text{ }^\circ\text{C}$ for the clean tube and $102.8 \text{ }^\circ\text{C}$ for the fouled tubes. These profiles are similar to these obtained for pine black liquor in Fig. 6.8.

Fig. 6.17 shows that with clean surfaces it is possible to obtain heat flux rates twice those of the fouled tubes with the assumed dirt resistance value.

The value of the overall heat transfer co-efficient decreases by a factor of 2 to 3 with the assumed value of dirt resistance $0.45 \text{ m}^2 \text{ }^\circ\text{K/KW}$ as demonstrated by the graph in Fig. 6.18.

The change in bamboo black liquor concentration in Fig. 6.19 shows that a final concentration of 27.8% and 15.7% can be obtained for the clean and fouled tubes respectively. This points out the need for relatively clean tube surfaces so that satisfactory evaporator capacity

levels can be maintained.

E. Evaporation of Bagasse Black Liquor and a Comparison with Bamboo Black Liquor Evaporation:

The computations for the evaporation of bagasse black liquors in the single effect LTV unit are identical to the procedure used earlier for pine and bamboo black liquors. The listing of computational procedure for bagasse black liquor is given in Appendix D-2 and is based on the same equipment geometry and process feed liquor conditions as used for the other black liquors in this investigation. Fig. 6.20-6.22 gives the profiles of temperature, overall heat transfer co-efficient and solid concentration along the tube length for both clean and fouled tube surfaces. The curves marked C-2 and D-2 denote clean and fouled evaporator tube ($R_D = 0.45 \text{ m}^2 \text{ }^\circ\text{K/KW}$) respectively for bagasse black liquor. For comparison purposes the corresponding results for bamboo black liquor are reproduced in these figures and the curves marked C-1 and D-1 represent clean and dirty tubes ($R_D = 0.45 \text{ m}^2 \text{ }^\circ\text{K/KW}$) respectively. Bagasse and Bamboo black liquors are very good examples for demonstrating the role of engineering properties of process liquors on the performance of evaporators.

The curve in Fig. 6.20 show that the temperature profile (C-1) and (D-1) for bamboo black liquor; and (C-2) and (D-2) for bagasse black liquor are similar. The length of the nonboiling region is 2.13 meter and 2.75 meter for the clean and fouled tubes respectively handling bagasse black liquor and the corresponding value for bamboo black liquor are 0.92 and 1.83 meter. This shows that the nonboiling section length is increased

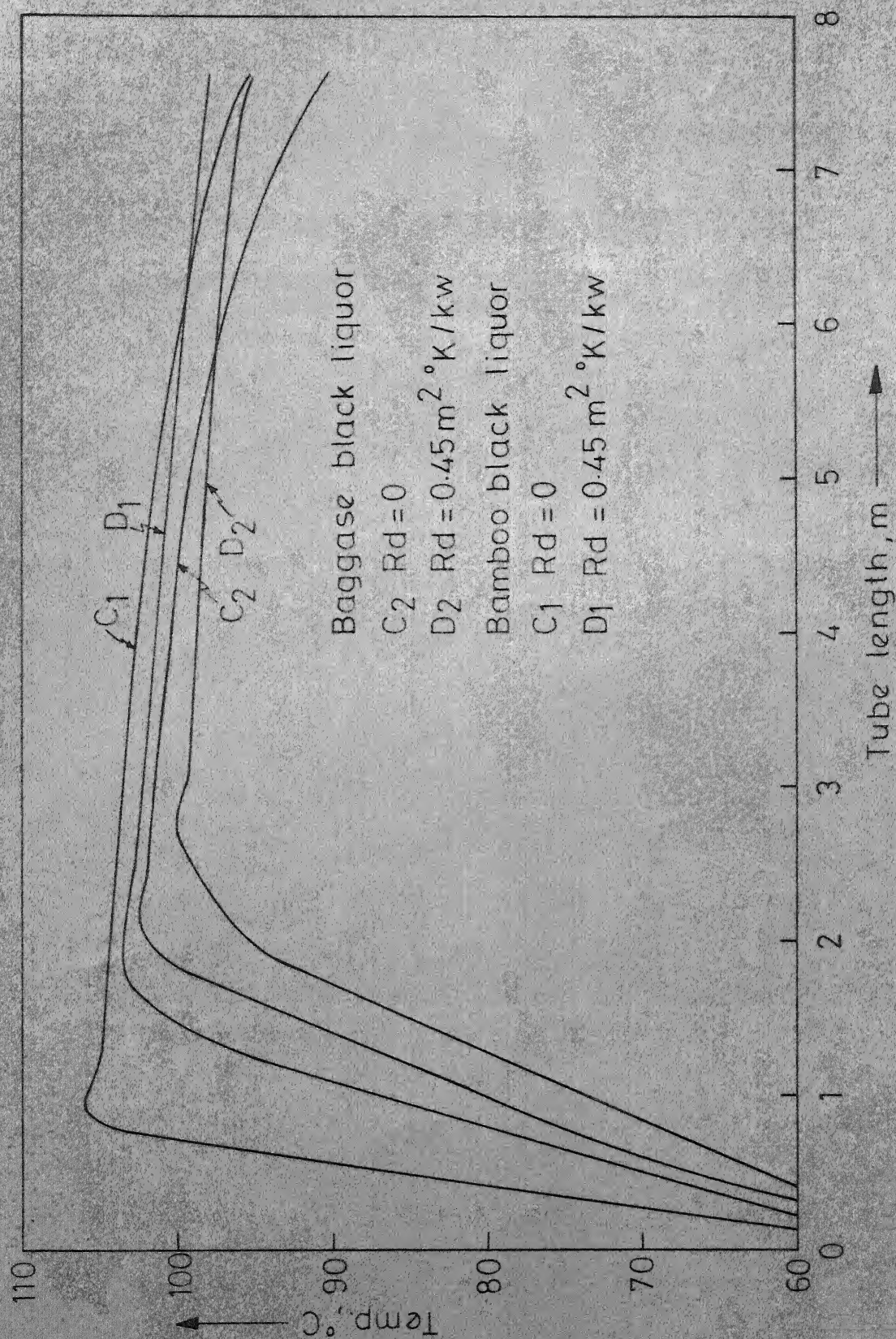


Fig.6.20 Variation of liquor temperature along tube length.

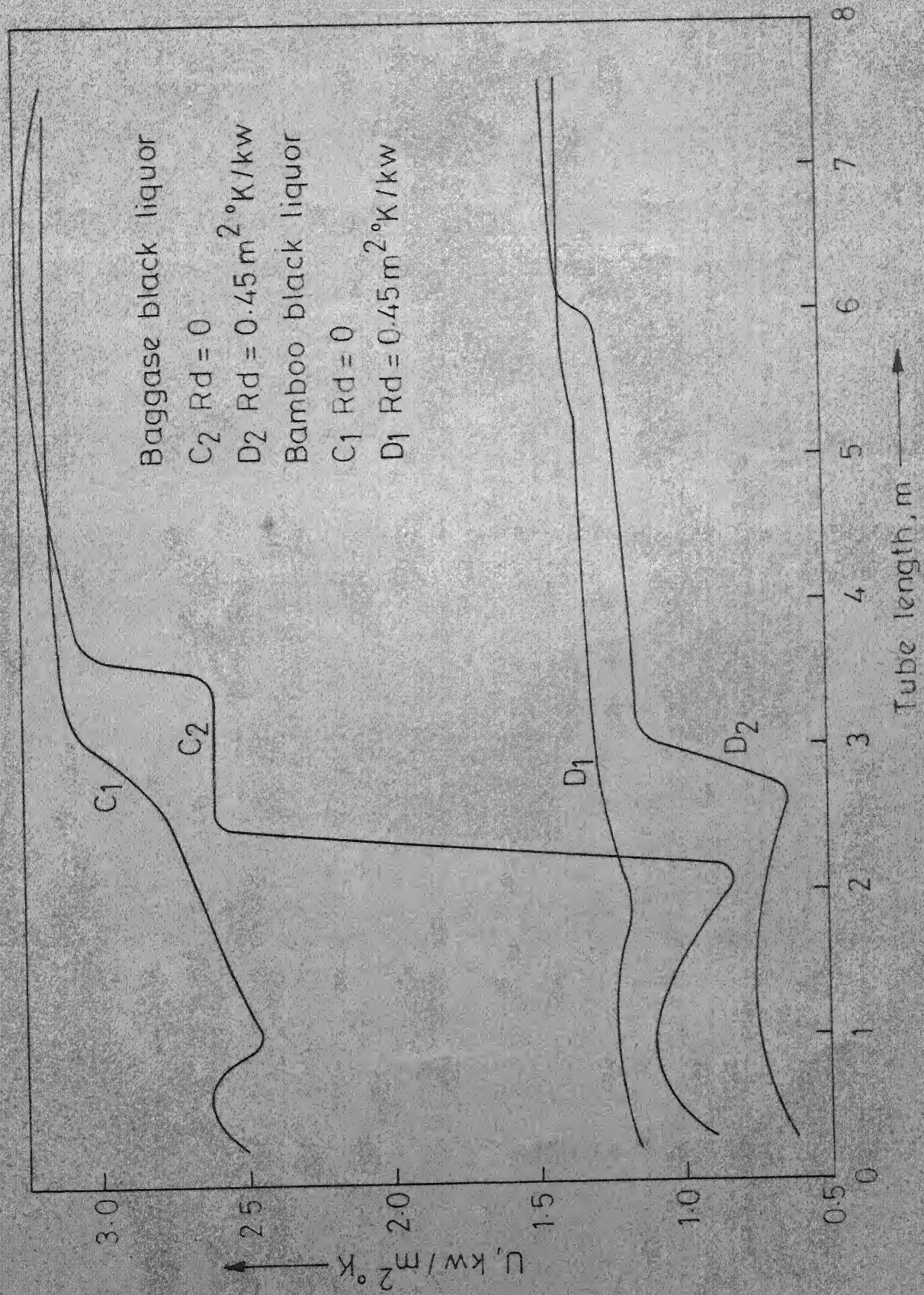


Fig.6.21-Overall heat transfer coefficient along tube length

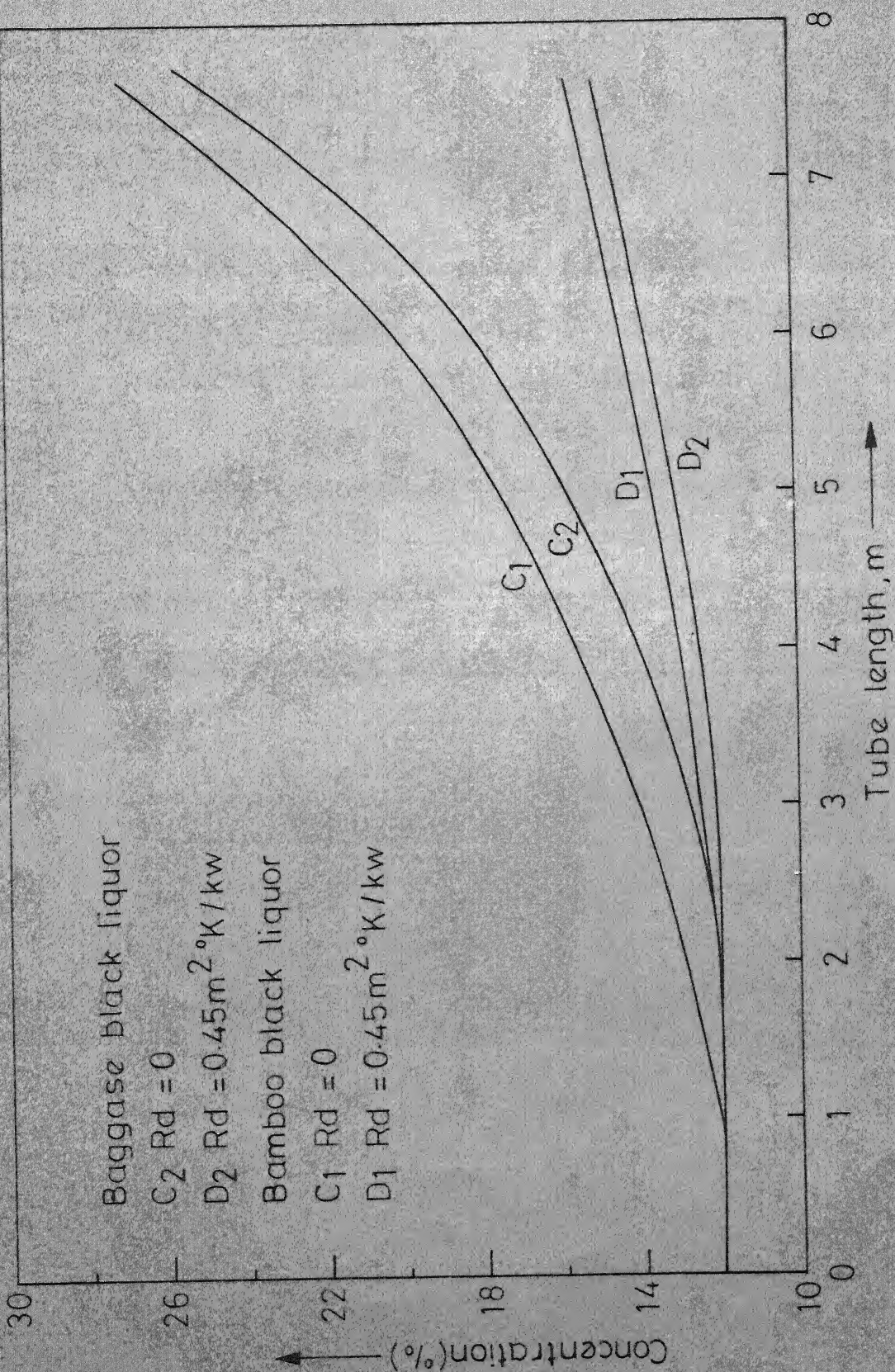


Fig.6.22-Variation of liquor concentration along tube length.

by a factor of 2.3 and 1.5 for clean and fouled surfaces respectively when changing over from bamboo to baggase black liquor. This can be attributed to the lower value of the overall heat transfer co-efficients for baggase black liquor as a result of its higher viscosity.

The curves in Fig. 6.21 show that the value of overall heat transfer co-efficient for bamboo black liquor in the nonboiling section is 2.5 times the value for baggase black liquor. This effect of liquor viscosity on the overall heat transfer co-efficient decreases the evaporation rate as may be observed by the curves in Fig. 6.22 (Curves C-1, C-2 and D-1, D-2). The viscosity of the Bamboo and baggase black liquor for the concentration range of Fig. 6.22 are given below in Table 6.4.

Table 6.4: Viscosity Range for Bamboo & Baggase Black Liquors in a Clean Tube

	Bamboo Black Liquor		Baggase Black Liquor	
	Inlet	Exit	Inlet	Exit
1. Nonboiling section-				
Temperature, °C	50.0	106.0	50.0	102.4
Solid Concentration, %	12.0	12.0	12.0	12.0
Viscosity, cp	0.9	0.45	4.7	1.9
2. Boiling section-				
Temperature, °C	106.0	95.3	102.4	86.2
Solid Concentration, %	12.0	27.8	12.0	25.9
Viscosity, cp	0.45	1.1	1.9	9.8

The results of the above comparison clearly shows that a given evaporator can be effectively used using preheated baggase black liquor to reduce the length of the nonboiling section. The role of wide variation in physical properties of black liquor can be appreciated in evaluating the performance of commercial plants utilizing multiple effect evaporators including auxilary heat recovery equipments such as liquid preheater, heat exchangers, condensate vaporizers, product vaporizers, vapor bleed etc. Evaporation plant in the chemical recovery system of kraft pulp mill usually consist of 4-7 effects with many of the heat recovery features listed above. The computational scheme for the single effect evaporation developed in this investigation can be applied to multiple effect evaporators which can be suitably modified to incorporate heat recovery features in the simulation programme.

CHAPTER 7SUMMARY

Simulation programme developed in this work based on slip model for two phase flow can give reliable values of the heat transfer rates in the long tube vertical evaporators. The model has been applied to black liquors over the concentration range of 12-28% dissolved solids and the values of the overall heat transfer co-efficients obtained agree very well with the experimental values reported in literature. The profiles of liquor temperature, concentration, heat flux and overall heat transfer co-efficients along the evaporator tube length are obtained for bamboo, baggase and pine black liquors and the influence of solution viscosity on the lengths of nonboiling, nucleate boiling and annular film regions in a given evaporation unit is illustrated. Reduction in evaporator capacity resulting from the dirt resistance to heat transfer from scale deposits can be predicted by this programme. The model can also be used to predict changes in evaporation capacity and the necessary auxiliary heat exchange equipment when the process liquor exhibits a large change in the viscosity.

NOMENCLATURE

A_L	Cross section of liquid stream, Sq ft.
A_G	Cross section of vapor stream, Sq.ft.
C_1	Cross sectional area of tube, Sq ft
C_L	Specific heat of liquid, Btu/lb °F
C_p	Specific heat of vapor at constant pressure, Btu/lb °F
D	Tube diameter, ft
g	Gravitational Acceleration, ft/sec ²
g_c	Conversion factor, 32.2 lb-ft/lb _f . sec ²
G_G	Mass flux of vapor stream, lb/sq ft. sec.
G_L	Mass flux of liquid stream, lb/sq ft. sec.
G	Mass flux of liquid-vapor stream, lb/sq ft. sec
J	Mechanical equivalent of heat, 778.17 ft-lb _f /Btu
P	Pressure, lb _f /sq ft
$\frac{dP}{dZ}$	Total pressure drop per unit length, lb _f /cu ft.
$\frac{dP_f}{dZ}$	Frictional pressure drop per unit length, lb _f /cu ft.
Q	Heat flux, Btu/hr. sq ft.
S	Solid concentration, Wt %
T	Temperature, °F
T_o	Reference temperature, °F
U	Velocity of homogeneous phase
V_G	Velocity of vapor stream, ft /sec

V_L	Velocity of liquid stream, ft/sec.
W_G	Mass flow rate of vapor, lb/sec
W_L	Mass flow rate of liquid, lb/sec
W_T	Total mass flow rate, lb/sec
W_S	Work done by the system on surrounding, Btu/lb-m
X	Quality of vapor-liquor mixture = W_G/W_T
Z	Tube length, ft
ρ_L	Density of liquid, lbm/cu ft.
ρ_G	Density of vapor, lbm/cu ft.
ρ	Density of vapor-liquid mixture, lbm/cu ft.
τ_w	Shear stress at the wall, lb _f /sq ft.
α	Void fraction
λ	Heat of vaporization, Btu/lbm

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APPENDIX A

TWO PHASE FRICTIONAL PRESSURE DROP

Lockhart & Martinelli⁽²⁴⁾ have analyzed four types of two phase flow mechanisms which exist during the simultaneous flow of a liquid and a gas or vapor: (1) Turbulent-Turbulent flow (2) Viscous-turbulent flow (3) Turbulent-viscous flow (4) Viscous-Viscous flow (flow of both liquid and gas may be viscous). This analysis is based on the following assumptions:

1. Static pressure drop for the liquid phase must equal the static pressure drop for the gaseous phase regardless of the flow pattern, as long as an appreciable radial static pressure difference does not exist.
2. The volume occupied by the liquid plus the volume occupied by the gas at any instance must equal the total volume of the pipe.
3. Flow mechanism transition criteria:

$Re_{gp} \geq 2000$ the gas phase is turbulent

$Re_{lp} \leq 1000$ flow of liquid is viscous

where $Re_{gp} = \frac{4 W_g}{\pi D_p \mu_g}$; $Re_{lp} = \frac{4 W_l}{\pi D_p \mu_l}$

Lockhart & Martinelli defined a parameter X (two phase flow modulus) for the different flow mechanisms given by Eq.A-1 to Eq.A-4.

$$X_{tt}^2 = \left(\frac{W_l}{W_g} \right)^{1.8} \left(\frac{\rho_g}{\rho_l} \right) \left(\frac{\mu_l}{\mu_g} \right)^{0.2} \quad (A-1)$$

$$X_{vt}^2 = \frac{C_1 \text{Re}_{gp}^{-0.8}}{C_g} \frac{W_1}{W_g} \frac{\rho_g}{\rho_1} \frac{\mu_1}{\mu_g} \quad (\text{A-2})$$

$$X_{tv}^2 = \frac{C_1 \text{Re}_{lp}^{0.8}}{C_g} \frac{W_1}{W_g} \frac{\rho_g}{\rho_1} \frac{\mu_1}{\mu_g} \quad (\text{A-3})$$

$$X_{vv}^2 = \frac{W_1}{W_g} \frac{\rho_g}{\rho_1} \frac{\mu_1}{\mu_g} \quad (\text{A-4})$$

The value of constants C_1 , C_g , n , m for use in equation A-1 to A-4 are given below:

Type of Flow	t-t	v-t	t-v	v-v
n	0.2	1.0	0.2	1.0
m	0.2	0.2	1.0	1.0
C_1	0.046	16.0	0.046	16.0
C_g	0.046	0.046	16.0	16.0

The following equations are obtained by least square techniques for the parameter $\phi_g \left[\left(\frac{\Delta P}{\Delta L} \right)_{TP} / \left(\frac{\Delta P}{\Delta L} \right)_g \right]$ and R_g (fraction of tube filled with the vapor) as a function of corresponding X , using curves given by Lockhart & Martinelli for ϕ_g and R_g .

$$\phi_{gtt} = \text{Exp} (1.54623 + 0.48518 \log X_{tt} + 0.044 (\log X_{tt})^2) \quad (\text{A-5})$$

$$\phi_{gvt} = \text{Exp} (1.36397 + 0.49115 \log X_{vt} + 0.05209 (\log X_{vt})^2) \quad (\text{A-6})$$

$$\begin{aligned} \phi_{gtv} = & \text{Exp} (1.35883 + 0.52153 \log X_{tv} + 0.05963 (\log X_{tv})^2 \\ & - 0.0013 (\log X_{tv})^3 - 0.0004 (\log X_{tv})^4) \end{aligned} \quad (\text{A-7})$$

$$\begin{aligned} \phi_{gvv} = & \text{Exp} (1.0705 + 0.5520 \log X_{vv} + 0.0881 (\log X_{vv})^2 \\ & - 0.0026 (\log X_{vv})^3 - 0.0012 (\log X_{vv})^4) \end{aligned} \quad (\text{A-8})$$

$$R_g = \text{Exp} (-0.2235 - 0.10046 \log X - 0.0404 (\log X)^2 - 0.0066 (\log X)^3) \quad (\text{A-9})$$

$$\phi \left[\left(\frac{\Delta P}{\Delta L} \right)_{TP} / \left(\frac{\Delta P}{\Delta L} \right)_L \right] \text{ and } R_1 \text{ (fraction of tube filled with}$$

the liquor) can be given by Eq. A-10 and A-11.

$$\phi_1 = \phi_g / X \quad (\text{A-10})$$

$$R_1 = 1 - R_g \quad (\text{A-11})$$

Parameter α (flow type modulus for liquid) and β (flow type modulus for vapor) are given by equation A-12 and A-13.

$$\alpha = R_1^4 \phi_{ltt}^{3.33} = R_1^2 \phi_{lvt}^2 = R_1^4 \phi_{ltv}^{3.33} = R_1^2 \phi_{lvv}^2 \quad (\text{A-12})$$

$$\beta = R_g^4 \phi_{gtt}^{3.33} = R_g^4 \phi_{gvt}^{3.33} = R_g^2 \phi_{gtv}^2 = R_g^2 \phi_{gvv}^2 \quad (\text{A-13})$$

D_1 (hydraulic diameter of liquid flow) and D_g (hydraulic diameter of gas flow) can be given by Eq. A-14 and A-15.

$$\left(\frac{D_p}{D_1}\right)^2 = \frac{\alpha}{R_1} \quad (\text{A-14})$$

$$\left(\frac{D_p}{D_g}\right)^2 = \frac{\beta}{R_g} \quad (\text{A-15})$$

Total frictional pressure drop can be calculated by Eq. A-16 by using equations A-1 to A-15.

$$\left(\frac{\Delta P}{\Delta L}\right)_{\text{fric}} = \frac{2 \left(\frac{4}{\pi}\right)^{2-n} C_1 \mu_1^n W_1^{(2-n)}}{D_p^{(5-n)} \rho_1 g} \alpha^{(n-2)} \left(\frac{D_p}{D_1}\right)^{(5-n)} \quad (\text{A-16})$$

For the case when $R_{\text{egp}} > 2100$ and $R_{\text{elp}} > 8000$, there is a correction by Davis⁽²⁵⁾. For this range Davis defined the parameter X given by Eq. A-17.

$$X = 0.19 \left(\frac{W_1}{W_g}\right)^{0.9} \left(\frac{\rho_g}{\rho_1}\right)^{0.5} \left(\frac{\mu_1}{\mu_g}\right)^{0.1} \left(\frac{V_m^2}{D_p g}\right)^{0.185} \quad (\text{A-17})$$

$$\text{Where } V_m = \left(\frac{W_1}{\rho_1} + \frac{W_g}{\rho_g}\right) / \frac{\pi D_p^2}{4}$$

Notations:

$(\Delta P / \Delta L)_{\text{fric}}$ = pressure drop per unit length due to friction, lb-f/cu ft

W_1, W_g = liquid flow rate and gas flow rate respectively, lb/sec.

D_p = Dia of pipe, ft. μ = viscosity, lb-m/sec.ft

ρ = Density lb-m/cu ft.

APPENDIX B

COMPARISON OF TWO METHODS TO CALCULATE NONBOILING SECTION HEAT TRANSFER

CO-EFFICIENT

For nonboiling section heat transfer co-efficient, two correlations can be used: equation given by Badger et.al.⁽⁶⁾ and equation of Martinelli & Boelter⁽⁴⁷⁾. Badger's correlation gives overall whereas Martinelli & Boelter's expression would give convective heat transfer co-efficient for inside liquid and includes the effect of natural convection. Badger's correlation-

$$U_{NB} = 0.06 \left(\frac{C \Delta T_{NB} f^2}{\mu_{NB}} \right)^{0.7} \quad (B-1)$$

Martinelli & Boelter's correlation-

$$Nu_{a.m.} = 1.75 F1 \left[\frac{GZ_{a.m.} + 0.0722 F2 \left(\frac{Gr Pr}{X/dw} \right)^{0.84}}{w} \right] \quad (B-2)$$

$$\text{where } F1 = \frac{3.14 Nu_{a.m.} / GZ_{a.m.}}{\log (2 + 3.14 Nu_{a.m.} / GZ_{a.m.}) (2 - 3.14 Nu_{a.m.} / GZ_{a.m.})} \quad (B-3)$$

$$F2 = 1 - .408 \sqrt{Nu_{a.m.} / GZ_{a.m.}} \quad \frac{Nu_{a.m.}}{GZ_{a.m.}} < 1.85 \quad (B-4)$$

$$= -1.63 \left(\sqrt{Nu_{a.m.} / GZ_{a.m.}} - 2 \right) \quad 1.85 < \frac{Nu_{a.m.}}{GZ_{a.m.}} < 2.0 \quad (B-5)$$

X = nonboiling section length, ft

Over all heat transfer co-efficient and nonboiling section length are calculated for process conditions of Badger data 7 (see Table 6.1) by

using eq. B-1 and eq. B-2; and are compared below:

Table B-1: Comparison of Results by Badger & Martinelli's Correlations

	Overall Heat Transfer Co-efficient Btu/hr sq ft °F	Nonboiling section length, ft
Badger et.al ⁽⁶⁾	90	6.32
Martinelli & Boelter ⁽⁴⁷⁾	84	6.53

As clear from Table B-1, both equations lead to nearly same results and for computational purposes it is simpler to use Badger correlation than Martinelli and Boelter expression; in the latter expression nonboiling section length is an implicit variable which will require a trial & error procedure for solution.

APPENDIX CTable C.1 - Antoine Equation Constants for Water

Temp. Range °C	Antoine Equation Constant		
	A	B	C
0 - 30	8.184254	1791.3	238.1
30 - 40	8.1393986	1767.262	236.29
40 - 50	8.0886767	1739.351	234.10
50 - 60	8.0464204	1715.429	232.14
60 - 70	8.0116295	1695.167	230.41
70 - 80	7.9845588	1678.948	228.97
80 - 90	7.9634288	1665.924	227.77
90 - 100	7.9483960	1656.390	226.86
100 - 150	7.9186968	1636.909	224.92

$$\log_{10} P = A - \frac{B}{T + C}$$

T = temperature, °C

P = pressure, mm of Hg

Table C.2 - Antoine Equation Constants for Water

Pressure Range mm of Hg	Antoini Equation Constants		
	A	B	C
5 - 30	8.184254	1791.3	238.1
30 - 35	8.1393986	1767.262	236.29
35 - 90	8.0886767	1739.351	234.10
90 - 150	8.0464202	1715.429	232.14
150 - 230	8.0116295	1695.167	230.41
230 - 340	7.9845588	1678.948	228.97
340 - 520	7.9634288	1665.924	227.77
520 - 760	7.948396	1656.39	226.86
760 -1500	7.9186968	1636.909	224.92

$$T = \frac{B}{A - \log_{10} P} - C$$

T = temperature, °C

P = pressure, mm of Hg

APPENDIX DCOMPUTER PROGRAMME LISTING

C1	Cross sectional area of the tube
TS	Steam temperature
DP	Internal dia. of tube
XL	Length of the tube
DZ	Differential length increment for computation
WL	Liquor mass flow rate
TI	Inlet temperature of the feed
S	Solid content in the feed
P	Pressure in the tube
WG	Vapor mass flow rate
BPT	Boiling point elevation
GPRLF	Frictional pressure gradient
UN	Overall heat transfer co-efficient
Q	Heat flux
DT	Temperature difference for the differential section
DPT	Pressure drop in the differential section
QT1	Heat transfered in the differential section
QT	Total heat transfered
X	Vapor quality
V	Homogeneous phase velocity
DEG	Density of vapor

XMUG	Viscosity of vapor
HV	Latent heat of vaporization for water
CP	Specific heat of vapor
DX	Vapor quality difference for the section
DS	Solid content change in the differential section
RGI	Void fraction at the inlet of the section
RGO	Void fraction at the outlet of the section
DV	Homogeneous phase velocity difference in the section
HTB	Heat transfered in the boiling section
US2	Average boiling heat transfer co-efficient
DEL	Density of liquor
CL	Specific heat of liquor
UN	Nonboiling section overall heat transfer co-efficient
XMUL	Viscosity of liquor
WT	Feed liquor mass flow rate
REG	Reynold no for vapor
REL	Reynold no for liquor
RL	Liquid volume fraction in the tube
KL	Thermal conductivity of liquor
ST	Surface tension of liquor
H	Inside tube heat transfer co-efficient for liquor
HS	(Wall + Steam side) Heat transfer co-efficient
U	Overall heat transfer co-efficient in the boiling section

DM	Mean diameter of the tube
DPO	Outside diameter of the tube
SK	Thermal conductivity of tube material
HO	Steam side heat transfer co-efficient
A,B,C	Antoine equation constants

£JOB CGG153,TIME008,PAGES040,NAME A.K.AGARWAL

£IBJOB

£IBFTC MAIN

C*****3

C*****SINGLE EVAPORATOR USING HOMOGENOUS MODEL*****

C*****PINWOOD BLACK LIQUOR IN A TUBE HAVING DIRT RESISTANCE*****

C*****3

COMMON C1,TS,DP

COMMON/LEVEL1/XL

COMMON/LEVEL2/DZ

READ 16,XL,DZ,DP

READ 17,WL,TS,T1,S,P

16 FORMAT(3F8.4)

17 FORMAT(5F10.3)

WG=0.

PRINT 101

101 FORMAT(-----INPUT-----)

PRINT 101,WL,TS,T1,S,P

103 FORMAT(* WL= *,F10.3,* LB=M/SECOND*,/* TS= *,F10.3,* DEGREE F *,/*

/ T=*,F10.2,* DEGREE F *,/* SOLID CONTENT= *,F10.3,* PERCENT *,/* P

/= *,F10.3,* LBF/SQ FT *)

PRINT 102

102 FORMAT(-----OUTPUT-----)

NE=1

C1=(3.14*DP**2.)/4.

T=T1

PRINT 1

1 FORMAT(-----EVAPORATOR NO -----)

CALL EVAP(WL,WG,S,T,P,NE)

STOP

END

£IBFTC EVAP

SUBROUTINE EVAP(WL,WG,S,T,P,NE)

C*****3

C*****SINGLE EVAPORATOR CALCULATIONS*****

C*****3

COMMON C,TS,DP

COMMON/LEVEL1/XL

COMMON/LEVEL2/DZ

XN=XL/DZ

DT=0.0

QT=0.0

N=0

CALL VPT(P,BPT)

C ****CHANGE CAPD FOR BOILING POINT ELEVATION****

BPT=BPT+.19*S*.75.

IF(T-BPT)37,35,35

37 IF(NE.EQ.2) GO TO 35

IF(NE.EQ.3) GO TO 35

1 DZ1=DZ

CALL FML(WL,T,S,GPRLF,U,Q,UN,DT,DT1,DPT,DZ1,TX,P)

QT1=Q*.14*DP*DZ1

QT=QT+QT1

PRINT 9

P=P+DPT


```

      DPF=GPRLF*DZ1
      PRINT 8,P,DPT,DPF
8      FORMAT(* PRESSURE= *,F16.8,* LB/SQ FT *,/** TOTAL PRESSURE DROP=
/*,F16.8,* LB/SQ FT *,/** FRICTIONAL PRESSURE DROP= *,F16.8,* LB/SQ
/ FT *)
9      FORMAT(110(1H-))
      PRINT 10,WL,S,T,U,UN,Q,QT1,QT
10     FORMAT(* MASS FLOW RATE OF BLACK LIQUOR= *,F16.4,* LB-M PER SECO
/ND *,/** SOLID CONTENT= *,F16.4,* PERCENT *,/** TEMPERATURE= *,F16
/ .4,* DEGREE FAHRENHEIT *,/** FLOW-VELOCITY= *,F16.4,* FT PER SECON
/D *,/** OVER-ALL HEAT TRANSFER CO-EFFICIENT= *,F16.4,* BTU/HR.SQ F
/T.DEG F *,/** HEAT FLUX= *,F16.4,* BTU/HR.SQ FT *,/** HEAT TRANSFE
/ RRED IN THE SECTION= *,F16.4,* BTU/HR *,/** TOTAL HEAT TRANSFERRED
/ UP TO THIS SECTION= *,F16.4,* BTU/HR *)
      N=N+1
      PRINT 14,TX
14     FORMAT(* BOILING POINT AT THIS PRESSURE= *,F16.4,* DEGREE F *)
      XL=XL-DZ1
      PRINT 18,XL,DZ1
18     FORMAT(* REMAINING HEIGHT= *,F16.2,* LENGTH INCREMENT= *,F16.3)
      PRINT 13,N
13     FORMAT(* COMPUTATION FOR SECTION *,I2,* HAS BEEN COMPLETED *)
      PRINT 9
      IF(T-TX)2,3,3
2      CONTINUE
      PRINT 11
11     FORMAT(* CALCULATION FOR NEXT LENGTH INCREMENT-- *)
      DT=0.0
      GO TO 1
3      PRINT 15
15     FORMAT(* CALCULATION FOR NONBOILING SECTION IS COMPLETE *)
35     PRINT 36
36     FORMAT(* THERE IS NO NONBOILING SECTION *)
C      BOILING SECTION NO-1
      XM=XL/DZ
      N=0
      WT=WL+WG
      X=WG/WT
      V=WT/(C1*62.43)
      TR=2.0
      DP1=-40.0
      DT=0.
      CALL PPL1,DT,DEG,DEG1,XMUG,HV,CP)
      DX=225.*.14*DP*DZ*(TS-T)/(3600.*WT*HV)
25     DV=0.
31     DT=0.
      DS=0.
      CALL EMC(WT,X,S,T,P,V,DX,DS,DT,DPP,DV,TP,Q,WL,WG,U,RGI,RGO)
      T=T+DT
      X=X+DX
      V=V+DV
      P=P+DPP
      S=S+DS
      QT1=Q*3.14*DP*DZ
      QT=QT+QT1

```



```

PRINT 9
PRINT 28,RGI
28 FORMAT(* VOID FRACTION AT INLET= *,F16.3)
PRINT 26,WL,WG,X,T,V,P,S,U,Q,QT1,QT,DPP
26 FORMAT(* WL= *,F16.4,* LB=M/SECOND *,/* WG= *,F16.4,* LBM/SECOND*,
/* X= *,F16.4,* T= *,F16.4,* DEGREE F *,/* V= *,F16.4,* FT/SECOND
/* *,/* PRESSURE= *,F16.4,* LBF/SQ.FT *,/* SOLID CONTENT= *,F16.4,*
/ PERCENT *,/* OVER ALL HEAT TRANSFER COEFFICIENT= *,F16.4,* BTU/H
/R.SQFT.DEGREE F *,/* HEAT FLUX= *,F16.4,* BTU/HR.SQFT *,/* HEAT TR
/ANSFERRED IN THE SECTION= *,F16.4,* BTU/HR *,/* TOTAL HEAT TRANSFER
/RRED= *,F16.4,* BTU/HR *,/* PRESSURE DROP= *,F16.4,* LBF/SQFT *)
PRINT 29,RGO
29 FORMAT(* VOID FRACTION AT OUTLET= *,F16.3)
PRINT 9
N=N+1
PRINT 13,N
IF(FLOAT(N)-XN)21,24,24
21 IF((XN-FLOAT(N))-1.)32,33,33
32 DZ=(XN-FLOAT(N))*DZ
33 IF(RGO-0.9) 34,30,30
34 IF(DPP-DPI) 22,23,23
22 IF(RGO-.8)23,30,30
30 TR=3.
PRINT 27
27 FORMAT(* REGION IS ANNULAR *)
23 DPI=DPP
GO TO 25
24 CONTINUE
RETURN
END
$IBFTC EML
SUBROUTINE EML(WL,T,S,GPRLF,U,Q,UN,DT,DT1,DPT,DZ,TX,P)
C*****
C*****ENERGY,MOMENTUM AND CONTINUITY EQUATIONS FOR SINGLE PHASE LIQUID***
C*****
COMMON C,TS,DP
5 CALL PRLT(WL,T,S,GPRLF,DPP,DZ)
CALL XD L(S,T,DEL)
CALL XCL(S,T,CL)
U=WL/(DEL*CL)
DPT=-DEL*DZ-1.0*GPRLF*DZ
DELT=TS-T
CALL XUN(TS,T,S,UN)
Q=UN*DELT
R=DP/2.0
A=Q*2.0*3.14*R*DZ/(CL*3600.0*DEL*U*CL)
B=DZ/(7.7817*100.0*CL)
DT1=A-B
IF(ABS(DT-DT1)-0.1)1,1,2
2 T=T+(DT1-DT)/2.0
DT=DT1
GO TO 5
1 T=T+(DT1+DT)/4.0
P=P+DPT
P=P*760.0/(144.0*14.7)

```



```

      TX=(1656.390/(7.9484-ALOG10(P)))-224.92
      P=P*144.0*14.7/760.0
      P=P-DPT
      TX=9.0*TX/5.0+32.0
      TX=TX+4.0
      IF(TX-T)7,6,6
7     IF((T-TX)-.5)6,6,8
8     DZ=DZ*.8
      T=(DT1+DT)/2.
      DT=0.
      GO TO 5
6     CONTINUE
      RETURN
      END
$IBFTC PRLT
      SUBROUTINE PRLT(WL,T,S,GPRLF,DPF,DZ)
C*****
C*****CALCULATES NONBOILING SECTION LIQUID FRICTIONAL PRESSURE DROP*****
C*****
      COMMON C1,TS,DP
      CALL MUL(S,T,XMUL)
      CALL XDEL(S,T,DEL)
      CALL MUL(S,TS,WMUL)
      V=WL/(DEL*C1)
      RE=DEL*V*DP/XMUL
      IF(RE-2100.0)1,1,2
1     F=16.0/RE
      F=F*(WMUL/XMUL)**0.38
      GO TO 3
2     A=4.0*ALOG10(2.0/(0.00015))
      F=(1.0/(A+2.28))**2.0
      F=F*(WMUL/XMUL)**0.17
3     DPF=-2.0*F*DZ*V**2.0*DEL/(DP*32.2)
      GPRLF=-DPF/DZ
      RETURN
      END
$IBFTC XUN
      SUBROUTINE XUN(TS,T,S,UN)
C*****
C*****NONBOILING SECTION OVERALL HEAT TRANSFER COEFFICIENT*****
C*****
      CALL XCL(S,T,CL)
      CALL MUL(S,T,XMUL)
      CALL XDEL(S,T,DEL)
      XMUL=3600.0*XMUL
      DELT=TS-T
      UN=0.06*(CL*DELT*DEL**2.0/XMUL)**0.7
      A=1./UN+.0025
      UN=1./A
      RETURN
      END
$IBFTC EMC
C*****
C*****HOMOGENEOUS MODEL FOR ENERGY,MOMENTUM AND CONTINUITY EQUATIONS*****
C*****

```



```

SUBROUTINE EMC(WT,X,S,T,P,V,DX,DS,DT,DPP,DV,TR,Q,WL,WG,U,RGI,RGO)
COMMON C1,TS,DP
COMMON/LEVEL2/DZ
10 CALL PRF(WT,X,T,S,DX,DT,DS,GPRF,RG,P)
CALL XDEL(S,T,DEL)
CALL PP(T,DT,DEG,DEG1,XMUG,HV,CP)
IF(X=0.)22,21,22
21 RGI=0.
GO TO 23
22 CALL HOLDUP(WT,X,T,S,RLI,RGI,P)
23 DET=(1.-RGI)*DEL+RGI*DEG
TO=T+DT
XO=X+DX
SO=S+DS
CALL HOLDUP(WT,XC,TO,SO,RLO,RGO,P)
CALL XDEL(SO,TO,DEL1)
DET1=(1.-RGO)*DEL1+RGO*DEG1
DETM=(DET+DET1)/2.
DPP=-DETM*DZ-1.*GPRF*DZ-DETM*V*DZ/32.2
A=(P+DPP)*760.0/(144.0*14.7)
TX=(1656.39/(7.9484-ALOG10(A)))-224.92
TX=9.0*TX/5.0+32.0
C ****CHANGE CARD FOR BOILING POINT ELEVATION****
BPE=.19*S*9./5.
DT=TX+BPE-T
CALL XCL(S,T,CL)
B=DETM*V*C1*3600./(3.14*DP*DZ)
1 C=(1.-X)*CL*DT+HV*DX+X*CP*DT+V*DZ/(32.2*778.17)+DZ/(778.17)
Q=B*C
IF(TR.EQ.2.) GO TO 2
CALL UAF(T,S,DT,DS,Q,RG,U,WT,X,DX)
GO TO 3
2 CALL UNB(T,S,P,DT,DS,DPP,Q,U,WT,X,DX)
3 Q=U*(TS-T=DT/2.)
DX1=(Q/B-(1.-X)*CL*DT-X*CP*DT-V*DZ/(32.2*778.17)-DZ/778.17)/HV
IF(ABS(DX-DX1) < .002)5,5,5
5 DX=(DX+DX1)/2.
13 FORMAT(60(1H*))
GO TO 1
6 DX=(DX+DX1)/2.
WG=WT*X
WL=WT-WG
DS=(S*WT*DX)/(WL-WT*DX)
TO=T+DT
XO=X+DX
SO=S+DS
PO=P+DPP
CALL HOLDUP(WT,XC,TO,SO,RLO,RGO,PO)
CALL PP(T,DT,DEG,DEG1,XMUG,HV,CP)
CALL XDEL(SO,TO,DEL1)
DET1=(1.-RGO)*DEL1+RGO*DEG1
DD=DET1-DET
DV1=-V*DD/DET
IF(ABS(DV-DV1) < .01)8,8,9
9 DV=DV1

```



```

      GO TO 10
8     DV=(DV+DV1)/2.
      RETURN
      END

```

```

SUBROUTINE PRF

```

```

C*****

```

```

C*****CALCULATION OF TWO PHASE FRICTIONAL PRESSURE DROP*****

```

```

C*****

```

```

      SUBROUTINE PRF(WT,X,T,S,DX,DT,DS,GPRF,PG,P)

```

```

      COMMON C1,TS,DP

```

```

      COMMON/LEVEL2/DZ

```

```

      T=T+DT/2.0

```

```

      X=X+DX/2.

```

```

      S=S+DS/2.0

```

```

      CALL MUL(S,T,XMUL)

```

```

      CALL XDEL(S,T,DEL)

```

```

      CALL PP(T,DT,DEG,DEG1,XMUG,HV,CP)

```

```

      WG=X*WT

```

```

      WL=WT-WG

```

```

      REG=4.0*WG/(3.14*DP*XMUG)

```

```

      REL=4.0*WL/(3.14*DP*XMUL)

```

```

      A=WL/WG

```

```

      B=DEG/DEL

```

```

      C=XMUL/XMUG

```

```

      IF(REG-2100.)3,1,1

```

```

1     IF(REL-8000.)3,2,2

```

```

2     VM=(WL/DEL+WG/DEG)/C1

```

```

      H=.19*A**.9*B**.5*C**.1*(VM**2./(DP**2.2))**.185

```

```

      GO TO 7

```

```

3     IF(REG-2000.)6,4,4

```

```

4     IF(REL-2000.)6,5,5

```

```

5     H=SQRT(A**1.8*B**C**.2)

```

```

7     PG=EXP(1.54623+0.48518*ALOG(H)+0.0444*(ALOG(H))**2)

```

```

      PL=PG/H

```

```

      CL=0.046

```

```

      XN=0.2

```

```

      D=ALOG(H)

```

```

      CALL HOLDUP(WT,X,T,S,PL,PG,P)

```

```

      ALPA=PL**4.*PL**3.33

```

```

      GO TO 14

```

```

6     IF(REG-2000.)10,8,8

```

```

8     IF(REL-2000.)9,10,10

```

```

9     CL=16.0

```

```

      CG=0.046

```

```

      XN=1.0

```

```

      H=SQRT((CL/CG)*REG**(-0.8)*A*B*C)

```

```

      D=ALOG(H)

```

```

      PG=EXP(1.36397+0.49115*D+0.05209*D**2)

```

```

      PL=PG/H

```

```

      CALL HOLDUP(WT,X,T,S,PL,PG,P)

```

```

      ALPA=PL**2.0*PL**2.0

```

```

      GO TO 14

```

```

10    IF(REG-2000.)11,13,13

```

```

11    IF(REL-2000.)13,12,12

```

```

12    CL=0.046

```



```

CG=16.0
XN=0.2
H=SQRT((CL/CG)*REL**(0.8)*A*B*C)
D=ALOG(H)
PG=EXP(1.35883+0.52153*D+0.05963*D**2-.0013*D**3-.0004*D**4.)
PL=PG/H
CALL HOLDUP(WT,X,T,S,RL,RG,P)
ALPA=RL**4.*PL**3.33
GO TO 14
13 H=SQRT(A*B*C)
CL=16.0
XN=1.0
D=ALOG(H)
PG=EXP(1.0705+.5520*D+.0881*D**2-.0026*D**3-.0012*D**4)
PL=PG/H
CALL HOLDUP(WT,X,T,S,RL,RG,P)
ALPA=RL**2.0*PL**2.0
14 F=SQRT(ALPA/RL)
F=2.*(4.*WL/3.14)**(2.-XN)*CL*XMUL**XN/(DP**(5.-XN)*DEL*32.2)
G=ALPA**(XN-2.)*F**(5.-XN)
GPRF=F*G
PRD=GPRF*DZ
T=T-DT/2.
X=X-DX/2.
S=S-DS/2.
RETURN
END
£IBFTC HOLDUP
SUBROUTINE HOLDUP(WT,X,T,S,RL,RG,P)
C*****
C*****CALCULATION OF VOID FRACTION IN THE TUBE*****
C*****
COMMON C1,DZ,TS,CP
CALL XDEL(S,T,DEL)
DT=0.
C=2.*DEL*X**2
B=DEG*(1.-X)**2-4.*DEL*X**2
A=2.*(D-G*(1.-X)**2-DEL*X**2)
CALL PP(T,DT,DEG,DEGT,XMUG,HV,CP)
D=SQRT(B**2-4.*A*C)
RG=(-B+D)/(2.*A)
RG1=(-B-D)/(2.*A)
IF(RG-1.)4,5,5
5 RG=RG1
GO TO 3
4 IF(RG-0.)2,7,3
2 RG=RG1
3 RL=1.-RG
RETURN
END
£IBFTC UNB
SUBROUTINE UNB(T,S,P,DT,DS,DPP,Q,U,WT,X,DX)
C*****
C*****CALCULATES OVERALL HEAT TRANSFER COEFFICIENT FOR NUCLEAT BOILING*****
C*****

```



```

COMMON C1, TS, DP
COMMON/LEVEL2/DZ
T=T+DT/2.
S=S+DS/2.
P=P+DPP/2.
CALL MUL(S,T,XMUL)
CALL XDEL(S,T,DEL)
CALL XCL(S,T,CL)
CALL KL(S,T,XKL)
CALL PP(T,DT,DEG,DEG1,XMUG,HV,CP)
ST=35./((453.6*32.2)
PRL=CL*XMUL*3600./XKL
WG=WT*(X+DX/2.)
WL=WT-WG
REL=4.0*WL/(DP*3.14*XMUL)
REG=4.0*WG/(DP*3.14*XMUG)
A=DEL/DEG
B=XMUG/XMUL
C=XKL/DP
H1=C*(1.3+39.*DP)*PRL**0.9*REL**0.23*REG**0.34*A**0.25*B
D=Q*DP/(HV*XMUL*3600.0)
E=P*DP/ST
F=A-1.0
H2=C*.22*PRL**.49*D**.69*E**.31*F**.33
IF(H1-H2)1,1,2
1 H=H2
GO TO 3
2 H=H1
3 CALL XHS(T,DT,H,C,HS)
G=1./H+1./HS
U=1./G
T=T-DT/2.
S=S-DS/2.
P=P-DPP/2.
RETURN
END

```

8IBFTC UAF

SUBROUTINE UAF(T,S,DT,DS,Q,RG,U,WT,X,DX)

C*****&
C*****CALCULATES OVERALL HEAT TRANSFER COEFFICIENT FOR ANNULAR FILM FLOW**&
C*****&

```

COMMON C1, TS, DP
COMMON/LEVEL2/DZ
T=T+DT/2.
S=S+DS/2.
CALL XCL(S,T,CL)
CALL XDEL(S,T,DEL)
CALL PP(T,DT,DEG,DEG1,XMUG,HV,CP)
WG=WT*(X+DX/2.)
ST=35./((453.6*32.2)
V=WG/(DEG*CL*RG)
A=.012*(V*SQRT(DP*DEG/(ST*32.2)))**0.5
B=CL*SQRT(DEL*ST*32.2/DP)
H=A*B*3600.0*4.

```

2 FORMAT(1* H FOR ANNULAR FLOW= *,F16.4)


```

CALL XHS(T,DT,H,C,HS)
G=1./H+1./HS
U=1./G
T=T-DT/2.
S=S-DS/2.
RETURN
END

```

£IBFTC XHS

SUBROUTINE XHS(T,DT,H,Q,HS)

```

C*****&
C*****CACCULATES(STEAM SIDE+WALL)HEAT TRANSFER COEFFICIENT*****&
C*****&

```

```

COMMON CI,TS,DP
COMMON/LEVEL1/XL
COMMON/LEVEL2/DZ
T=T+DT/2.
DX=0.0083
DM=0.1743
DPO=0.1820
SK=26.0
R=1./H+DX*DP/(SK*DM)+.0025
R=1./R
TW=T+Q/R
TF=TS-3.*(TS-TW)/4.
XMUW=.00014
XKW=0.394/3600.0
DEW=0.932615*62.43
CALL PP(TS,DT,DEG,DEG1,XMUG,HV,CP)
HO=1.13*(XKW**3.*DEW**2.*32.2*HV/(XL*XMUG*(TS-TW))**.25
HO=HO*3600.0
A=DX*DP/(SK*DM)+CP/(HO*DPO)+.0025
HS=1./A
T=T-DT/2.
RETURN
END

```

£IBFTC PP

SUBROUTINE PP(T,DT,DEG,DEG1,XMUG,HV,CP)

```

C*****&
C*****PHYSICAL PROPERTIES OF WATER VAPOR*****&
C*****&

```

```

T=5.0*(T-32.0)/9.0+273.0
CP=(7.7+.00045*T+.2521E-05*T**2.-.8587E-09*T**3.)/18.0
T=T-273.0
XMUG=(91.9+0.351*T)*.672E-07
DEG=21495.61221632-400.15672832*T+2.5991631*T**2.-.00579602*T**3.
DEG=62.43/DEG
HV=2799.1422976-18.2096504*T+.20634452*T**2.-.00142704*T**3.
/+1.00000353*T**4.
HV=HV*453.5924/1054.8
T=9.0*T/5.0+32.0
T1=5.0*(T+DT)-32.0)/9.0
DEG1=21493.61221632-400.15672832*T1+2.5991631*T1**2.-.00579602*
/T1**3.
DEG1=62.43/DEG1
RETURN

```



```

      END
      EIBFTC KL
      SUBROUTINE KL(S,T,XKL)
C*****
C ****PINE-WOOD BLACK LIQUOR THERMAL CONDUCTIVITY****
C ****BAMBOO*****
C*****
      T=5.0*(T-32.0)/9.0
      XKL=0.53354269-.03425*S+ (.0011355+.0000209*S)*T
      XKL=XKL/1.49
      T=9.0*T/5.0+32.0
      RETURN
      END
      EIBFTC MUL
      SUBROUTINE MUL(S,T,XMUL)
C*****
C ****PINE-WOOD BLACK LIQUOR VISCOSITY****
C*****
      T=5.0*(T-32.0)/9.0
      A=0.4717-0.02472*T+0.7059*10.0**(-5.0)*T**(2.0)
      B=0.06973-0.5452*10.0**(-3.0)*T+0.1656*10.0**(-5.0)*T**(2.0)
      C=0.002046+0.3183*10.0**(-4.0)*T-0.9761*10.0**(-7.0)*T**(2.0)
      D=0.5793*10.0**(-4.0)-0.6129*10.0**(-6.0)*T+0.1837*10.0**(-8.0)*T*
/*(2.0)
      XMUL=EXP(A+B*S+C*S**(2.0)+D*S**(3.0))
      XMUL=XMUL*0.000672
      T=9.0*T/5.0+32.0
      RETURN
      END
      EIBFTC XDEL
      SUBROUTINE XDEL(S,T,DEL)
C*****
C ****PINE-WOOD BLACK LIQUOR DENSITY****
C*****
      T=5.0*(T-32.0)/9.0
      DEL=1007.0-0.495*T+6.0*S
      DEL=DEL*0.06243
      T=9.0*T/5.0+32.0
      RETURN
      END
      EIBFTC XCL
      SUBROUTINE XCL(S,T,CL)
C*****
C ****PINE-WOOD BLACK LIQUOR SPECIFIC HEAT****
C*****
      T=5.0*(T-32.0)/9.0
      CL=(1.0-0.0054*S)*4187.0
      CL=CL*0.000239
      T=9.0*T/5.0+32.0
      RETURN
      END

```

ENTRY

25.	.50	.166			
.185	274.	122.	12.	2400.	

EJOB CCG153, TIME008, PAGES020, NAME A.K.AGARWAL

LIBJOB

LIBFC MAIN

C*****

C*****SINGLE EVAPORATOR USING SLIP MODEL

C*****BAGGASE BLACK LIQUOR IN A CLEAN TUBE****

C*****

COMMON C1, TS, DP

COMMON/LEVEL1/XL

COMMON/LEVEL2/DZ

C SINGLE EVAPORATOR-BAGGASE-NO RESISTENCE

READ 16, XL, DZ, DP

READ 17, WL, TS, TI, S, P

16 FORMAT(3F8.4)

17 FORMAT(5F10.3)

WG=0.

PRINT 110

110 FORMAT(1H1)

PRINT 101

101 FORMAT('-----INPUT-----')

PRINT 103, WL, TS, TI, S, P

103 FORMAT('WL= ', F10.3, ' LB-M/SECOND', /, ' TS= ', F10.3, ' DEGREE F ', /,

/ ' T= ', F10.2, ' DEGREE F ', /, ' SOLID CONTENT= ', F10.3, ' PERCENT ', /, ' P

/= ', F10.3, ' -BF/SQ FT ')

PRINT 102

102 FORMAT('-----OUTPUT-----')

NE=1

C1=(7.1*DP**2.)/4.

T=TI

PRINT 1

1 FORMAT('-----EVAPORATOR NO 1-----')

CALL EVAP(WL, WG, S, T, P, NE)

STOP

END

LIBFC EVAP

SUBROUTINE EVAP(WL, WG, S, T, P, NE)

C*****

C*****SINGLE EVAPORATOR CALCULATIONS****

C*****

COMMON C1, TS, DP

COMMON/LEVEL1/XL

COMMON/LEVEL2/DZ

K=2

KN=XL/DZ

DT=0.0

QT=0.0

N=0

P=P*760.0/(144.0*14.7)

BPT=(1656.390/(7.9484-ALOG10(P)))-224.92

P=P*144.0*14.7/760.0

BPT=9.*BPT/5.+32.

CHANGE CARD FOR BOILING POINT ELEVATION

BPT=BPT+.19*5*9./5.

IF(T-BPT)37,35,35

37 IF(NE.EQ.2) GO TO 35

IF(NE.EQ.3) GO TO 35


```

1  DZ1=DZ
   CALL EML(WL,T,S,GPRLF,U,Q,UN,DT,DT1,DPT,DZ1,TX,P)
   QT1=Q*3.14*DP*DZ1
   QT=QT+QT1
   PRINT 9
   P=P+DPT
   DPF=-GPRLF*DZ1
   PRINT 8,P,DPT,DPF
8  FORMAT(* PRESSURE= *,F16.8,* LB/SQ FT *,//* TOTAL PRESSURE DROP=
  /*,F16.8,* LB/SQ FT *,//* FRICTIONAL PRESSURE DROP= *,F16.8,* LB/SQ
  / FT *)
9  FORMAT(110(LH=))
   PRINT 10,WL,S,T,U,UN,Q,QT1,QT
10  FORMAT(* MASS FLOW RATE OF BLACK LIQUOR= *,F16.4,* LB-M PER SECO
  /ND *,//* SOLID CONTENT= *,F16.4,* PERCENT *,//* TEMPERATURE= *,F16
  /4,* DEGREE FAHRENHEIT *,//* FLOW-VELOCITY= *,F16.4,* FT PER SECON
  /D *,//* OVER-ALL HEAT TRANSFER CO-EFFICIENT= *,F16.4,* BTU/HR.SQ F
  /T.DEG F *,//* HEAT FLUX= *,F16.4,* BTU/HR.SQ FT *,//* HEAT TRANSFER
  /RED IN THE SECTION= *,F16.4,* BTU/HR *,//* TOTAL HEAT TRANSFERRED
  / UP TO THIS SECTION= *,F16.4,* BTU/HR *)
   N=N+1
   PRINT 14,TX
14  FORMAT(* BOILING POINT AT THIS PRESSURE= *,F16.4,* DEGREE F *)
   XL=XL-DZ1
   PRINT 18,XL,DZ1
18  FORMAT(* REMAINING HEIGHT= *,F16.2,* LENGTH INCREMENT= *,F16.3)
   PRINT 13,N
13  FORMAT(* COMPUTATION FOR SECTION *,I3,* HAS BEEN COMPLETED *)
   PRINT 9
   IF(N.LT.K) GO TO 51
   IF(N.EQ.K) GO TO 52
52  PRINT 53
53  FORMAT(1H1)
   K=K+2
51  CONTINUE
   IF(T-TX)/2,3,3
2  CONTINUE
   PRINT 11
11  FORMAT(* CALCULATION FOR NEXT LENGTH INCREMENT= *)
   DT=0.0
   GO TO 1
3  PRINT 15
15  FORMAT(* CALCULATION FOR NONBOILING SECTION IS COMPLETE *)
35  PRINT 36
36  FORMAT(* THERE IS NO NONBOILING SECTION *)
   BOILING SECTION NO=1
   K=3
   XN=XL/DZ
   N=0
   WT=WL+WG
   X=WG/WT
   V=WT/(C)*62.431
   TR=2.0
   DP1=-40.0
   DT=0.
   CALL PP(T,DT,DEG,DEGT,XMUG,HV,CP)

```



```

DX=225.*3.14*DP*DZ*(TS-T)/(3600.*WT*HV)
25 DV=0.
31 DT=0.
DS=0.
CALL EMC(WT,X,S,T,P,V,DX,DS,DT,DPP,DV,TR,Q,WL,WG,U,RGI,RG3)
T=T+DT
X=X+DX
V=V+DV
P=P+DPP
S=S+DS
CALL XDEL(S,T,DEL)
V=WL/(DEL*CI*(1.-RG3))
QT1=Q*3.14*DP*DZ
QT=QT+QT1
PRINT 9
PRINT 28,RGI
28 FORMAT(* VOID FRACTION AT INLET= *,F16.3)
PRINT 26,WL,WG,X,T,V,P,S,U,Q,QT1,QT,DPP
26 FORMAT(* WL= *,F16.4,* LB-M/SECOND *,/* WG= *,F16.4,* LBM/SECOND*,
/* X= *,F16.4,* T= *,F16.4,* DEGREE F *,/* V= *,F16.4,* FT/SECOND
/* PRESSURE= *,F16.4,* LBF/SQ.FT *,/* SOLID CONTENT= *,F16.4,*
/ PERCENT *,/* OVER ALL HEAT TRANSFER CO-EFFICIENT= *,F16.4,* BTU/H
/R.SQFT.DEGREE F *,/* HEAT FLUX= *,F16.4,* BTU/HR.SQFT *,/* HEAT TR
/ANSFERRED IN THE SECTION= *,F16.4,* BTU/HR *,/* TOTAL HEAT TRANSFER
/RED= *,F16.4,* BTU/HR *,/* PRESSURE DROP= *,F16.4,* LBF/SQFT *)
PRINT 29,RG3
29 FORMAT(* VOID FRATION AT OUTLET= *,F16.3)
PRINT 9
N=N+1
PRINT 13,N
IF(N.LT.K) GO TO 54
IF(N.EQ.K) GO TO 55
55 PRINT 33
K=K+3
54 CONTINUE
IF(FLOAT(N)-IN)21,24,24
21 IF((XN-FLOAT(N))+1.)32,33,33
32 DZ=(XN-FLOAT(N))*DZ
33 IF(RG3-0.9) 34,30,30
34 IF(DPP-DP1) 22,23,23
22 IF(RG3-.8)23,30,30
30 TR=3.
PRINT 27
27 FORMAT(* REGION IS ANNULAR *)
23 DP1=DPP
GO TO 25
24 CONTINUE
RETURN
END
$IBFIC EML
SUBROUTINE EML(WL,T,S,GRLEF,U,Q,UN,DT,DT1,DPT,DZ,TX,P)
*****
C*****ENERGY,MOMENTUM AND CONTINUITY EQUATIONS FOR SINGLE PHASE LIQUID***
C*****
COMMON C1,TS,DP
5 CALL PRLT(WL,T,S,GRLEF,DPP,DZ)

```



```

CALL XDEL(S,T,DEL)
CALL XCL(S,T,CL)
U=WL/(DEL*CI)
DPT=-DEL*DZ-1.0*GPRLF*DZ
DELT=TS-T
CALL XUN(TS,T,S,UN)
Q=UN*DELT
R=DP/2.0
A=Q*2.0*3.14*R*DZ/(CL*7600.0*DEL*U*CI)
B=DZ/(17.7817*100.0*CL)
DT1=A-B
IF(ABS(DT-DT1)-0.1)1,1,2
2 T=T+(DT1-DT)/2.0
DT=DT1
GO TO 5
1 T=T+(DT1+DT)/4.0
P=P+DPT
P=P*760.0/(164.0*16.7)
TX=(1656.390/(7.9*84-ALOG10(P)))-224.92
P=P*164.0*16.7/760.0
P=P-DPT
TX=9.0*T/5.0+52.0
TX=TX+4.0
IF(TX-T)7,6,5
7 IF((T-TX)-.5)6,6,8
8 DZ=DZ+.8
T=T-(DT1+DT)/2.
DT=0.
GO TO 5
6 CONTINUE
RETURN
END
CIBFTC PRLT
SUBROUTINE PRLT(WL,T,S,GPRLF,DPF,DZ)
C*****
C*****CALCULATES NONBOILING SECTION LIQUID FRICTIONAL PRESSURE DROP*****
C*****
COMMON C,TS,DP
CALL MUL(S,T,XMUL)
CALL XDEL(S,T,DEL)
CALL MUL(S,TS,WMUL)
V=WL/(DEL*CI)
RE=DEL*V*DP/XMUL
IF(RE-2.00-0)1,1,2
1 F=16.0/RE
F=F*(WMUL/XMUL)**0.78
GO TO 3
2 A=4.0*ALOG10(2.0/0.0005)
F=(1.0/(A+2.28))**2.0
F=F*(WMUL/XMUL)**0.17
3 DPF=-2.0*F*DZ*V**(2.0)*DEL/(DP*32.2)
GPRLF=-DPF/DZ
RETURN
END
CIBFTC XUN
SUBROUTINE XUN(TS,T,S,UN)

```



```

C*****
C****NONBOILING SECTION OVERALL HEAT TRANSFER COEFFICIENT****
C*****
  CALL XCL(S,T,CL)
  CALL MUL(S,T,XMUL)
  CALL XDEL(S,T,DEL)
  XMUL=3600.0*XMUL
  DELT=TS-T
  UN=0.06*(CL*DELT*DEL**2.0/XMUL)**0.7
  A=1./UN
  UN=1./A
  RETURN
  END
LIBFTC EMC
  SUBROUTINE EMC(WT,X,S,T,P,V,DX,DS,DT,DPP,DV,TR,Q,WL,WG,U,RGI,RGO)
C*****
C****SLIP MODEL FOR ENERGY,MOMENTUM AND CONTINUITY EQUATIONS****
C*****
  COMMON C,TS,DP
  COMMON/L/VELP/DZ
  10 CALL PRFIWT,X,T,S,DX,DT,DS,GPRF,RG,P)
  CALL XDEL(S,T,DEL)
  CALL PP(T,DT,DEG,DEG1,XMUG,HV,CP)
  IF(X=0.)22,21,22
  21 RGI=0.
  GO TO 23
  22 CALL HOLDUP(WT,X,T,S,RLI,RGI,P)
  23 DET=(1.-RGI)*DEL+RGI*DEG
  TU=T+DT
  XU=X+DX
  SU=S+DS
  CALL HOLDUP(WT,XU,TU,SU,RLU,RGU,P)
  CALL XDEL(SU,TU,DEL1)
  DET1=(1.-RGU)*DEL1+RGU*DEG1
  DETM=(DET+DET1)/2.
  ACI=(1.-X)**2/(DEL*(1.-RGI))+X**2/(DEG*RGI)
  ACO=(1.-X)**2/(DEL1*(1.-RGU))+XU**2/(DEG1*RGU)
  AM=ACO-ACI
  DPP=DETM*DZ-GPRF*DZ-WT**2*AM/(C1**2)
  A=(P*DPP)*750.0/(144.0*4.7)
  TX=(1656.39/(7.9484-ALOG10(A)))-225.92
  TX=9.0*T/5.0+22.0
C  ****CHANGE CARD FOR BOILING POINT ELEVATION****
  BPE=.19*3*9.75.
  DT=TX+BP-T
  CALL XCL(S,T,CL)
  AE1=X**2/(DEG**2*RG1**2)+(1.-X)**2/(DEL**2*(1.-RGI)**2)
  AE0=XU**2/(DEG1**2*RGU**2)+(1.-XU)**2/(DEL1**2*(1.-RGU)**2)
  AE=AE0-A
  B=WT*3600./(1.-X)*DP*DZ)
  C=(1.-X)*CL*DT+HV*DX+X*CP*DT+AE*WT**2/(2.*32.2*778.17*C1**2)+DZ/77
  /P.17
  Q=B*C
  IF(TR.EQ.2.) GO TO 2
  CALL UAFIT,S,DT,DS,2,RG,U,WT,X,DX)
  GO TO 3

```



```

2 CALL UNBIT(S,P,DT,DS,DPP,Q,U,WT,X,DX)
3 Q=U*(TS-T+DT/2.)
DX)=(Q/B-(1.-X)*CL*DT-X*CP*DT-AE*WT**2/(2.*32.2*778.17*CL**2)-DZ/7
/78.17)/HV
IF (ABS(DX-DX1)-.0005)6,6,5
5 DX=(DX+DX1)/2.
11 FORMAT(60(1H*))
GO TO 10
6 DX=(DX+DX1)/2.
WG=WT*X
WL=WT-WG
DS=(S-WT*DX)/(WL-WT*DX)
RETURN
END

```

LIBFTC PRF

SUBROUTINE PRF(WT,X,T,S,DX,DT,DS,GPRF,RG,P)

C*****
C*****CALCULATION OF TWO PHASE FRICTIONAL PRESSURE DROP*****
C*****

```

COMMON C,TS,DP
COMMON/LEVEL2/DZ
T=T+DT/2.0
X=X+DX/2.
S=S+DS/2.0
CALL MUL(S,T,XMUL)
CALL XDEL(S,T,DEL)
CALL PR(T,DT,DEB,DEG,XMUG,HV,CP)
WG=X*WT
WL=WT-WG
REG=4.0*WG/(3.14*DP*XMUG)
REL=4.0*WL/(3.14*DP*XMUL)
A=WL/WG
B=DEG/DEL
C=XMUL/XMUG
IF (REG-2000.)3,1,1
1 IF (REL-2000.)3,2,2
2 VM=(WL/D+L+WG/DEB)/C
H=.19*A**-.9*B**-.9*C**1.7*(VM**2/(DP*32.2))**0.185
GO TO 7
3 IF (REG-2000.)6,6,6
4 IF (REL-2000.)6,5,5
5 H=SQRT(A**1.8*B**C**0.2)
7 PG=EXP(1.54613+0.48718*ALOG(H)+0.0444*(ALOG(H))**2)
PL=PG/H
CL=0.046
XN=0.2
D=ALOG(H)
CALL HOLOUP(WT,X,T,S,CL,REG,P)
ALPA=RL**4.*PL**3.33
GO TO 14
6 IF (REG-2000.)10,8,8
8 IF (REL-2000.)9,10,10
9 CL=16.0
CG=0.046
XN=1.0
H=SQRT((CL/CG)*REG**(-0.8)*A*B*C)

```



```

D=ALOG(H)
PG=EXP(1.36397+0.49115*D+0.05209*D**2)
PL=PG/H
CALL HOLDUP(WT,X,T,S,RL,RG,P)
ALPA=RL**2.0*PL**2.0
GO TO 14
10 IF(REG-2000.)11,13,13
11 IF(REL-2000.)13,12,12
12 CL=0.046
CG=16.0
XN=0.2
H=SQRT((CL/CG)*REL**(.0.8)*A*B*C)
D=ALOG(H)
PG=EXP(1.55883+0.52153*D+0.05963*D**2-.0013*D**3-.0004*D**4.)
PL=PG/H
CALL HOLDUP(WT,X,T,S,RL,RG,P)
ALPA=RL**4.*PL**3.33
GO TO 14
13 H=SQRT(A*B*C)
CL=16.0
XN=1.0
D=ALOG(H)
PG=EXP(1.0705+.5520*D+.0881*D**2-.0025*D**3-.0012*D**4)
PL=PG/H
CALL HOLDUP(WT,X,T,S,RL,RG,P)
ALPA=RL**2.0*PL**2.0
14 E=SQRT(ALPA/RL)
F=2.*(4.*WL/2.34)**(2.*XN)*CL*XMUL**XN/(DP**((5.-XN)*DEL**2.2))
G=ALPA**((XN-2.)*E**((5.-XN))
GPRF=F*G
PRD=GPRF*DZ
T=T-DT/2.
X=X-DX/2.
S=S-DS/2.
RETURN
END
&IBFTC HOLDUP
SUBROUTINE HOLDUP(WT,X,T,S,RL,RG,P)
C*****
C*****CALCULATION OF VOID FRACTION IN THE TUBE*****
C*****
COMMON C,DZ,TS,DP
CALL XDEL(S,T,DEL)
DT=0.
CALL PP(T,DT,DEG,D*GL,XMUG,HV,CP)
A=-2.*(D*GL*(1.-X)**2-DEL*X**2)
B=DEG*(1.-X)**2-4.*DEL*X**2
C=2.*DEL*X**2
D=SQRT(B**2-4.*A*C)
RG=(-B+D)/(2.*A)
RG1=(-B-D)/(2.*A)
IF(RG-1.)4,5,5
5 RG=RG1
GO TO 3
4 IF(RG-0.)2,3,3
2 RG=RG1

```



```

3  PL=1.-RG
RETURN
END
&IBFTC UNB
SUBROUTINE UNB(T,S,P,DT,DS,DPP,Q,U,WT,X,DX)
C*****
C*****CALCULATES OVERALL HEAT TRANSFER COEFFICIENT FOR NUCLEAT BOILING**
C*****
COMMON CL,TS,DP
COMMON/LEVEL2/DZ
T=T+DT/2.
S=S+DS/2.
P=P+DPP/2.
CALL MUL(S,T,XMUL)
CALL XDEL(S,T,DEL)
CALL XCL(S,T,CL)
CALL KL(S,T,XKL)
CALL PP(T,DT,DEG,DEG1,XMUG,HV,CP)
ST=35.7/(458.8*32.2)
PRL=CL*XMUL*3600./XKL
WG=WT*(X+DX/2.)
WL=WT-WG
REL=4.0*WL/(DP*3.14*XMUL)
REG=4.0*WG/(DP*3.14*XMUG)
A=DEL/DEG
B=XMUG/XMUL
C=XKL/DP
H1=C*(1.7+39.*DP)*PRL**0.9*REL**0.23*REG**0.6*A**0.25*B
D=Q*DP/(HV*XMUL*3600.0)
E=P*DP/ST
F=A-1.0
H2=C*.225*PRL**0.69*D**0.69*E**0.31*F**0.33
IF(H1-H2)1,1,2
1  H=H2
GO TO 3
2  H=H1
CALL XHS(T,DT,H,Q,HS)
G=1./H+1./HS
U=1./G
T=T-DT/2.
S=S-DS/2.
P=P-DPP/2.
RETURN
END
&IBFTC UAF
SUBROUTINE UAF(T,S,DT,DS,Q,RG,U,WT,X,DX)
C*****
C*****CALCULATES OVERALL HEAT TRANSFER COEFFICIENT FOR ANNULAR FILM FLOW**
C*****
COMMON CL,TS,DP
COMMON/LEVEL2/DZ
T=T+DT/2.
S=S+DS/2.
CALL XCL(S,T,CL)
CALL XDEL(S,T,DEL)
CALL PP(T,DT,DEG,DEG1,XMUG,HV,CP)

```



```

WG=WT*(X+DX/2.)
ST=5.7/(453.5*32.2)
V=WG/(DEG*CL*RG)
A=.012*(V*SQR(DP*DEG/(ST*32.2)))**0.5
B=CL*SQR(DEG*ST*32.2/DP)
H=A*B*3600.0*4.
2  FORMAT(* H FJR ANNULAR FLOW= *,F15.4)
CALL XHS(T,DT,H,Q,HS)
G=1./H+1./HS
U=1./G
T=T-DT/2.
S=S-DS/2.
RETURN
END
$IBFTC KHS
SUBROUTINE XHS(T,DT,H,Q,HS)
C*****
C*****CALCULATES (STEAM SIDE+WALL) HEAT TRANSFER COEFFICIENT*****
C*****
COMMON CL,TS,DP
COMMON/LEVEL1/XL
COMMON/LEVEL2/DZ
T=T+DT/2.
DX=0.0083
DM=0.1743
DPO=0.1820
SK=26.0
R=1./H+DK*DP/(SK*DM)
R=1./R
TW=T+Q/R
TF=TS-3.*(TS-TW)/4.
TF=5.*(T-32.)/9.
XMUG=.00014
XKW=.394/3600.
DEW=.912615*62.43
CALL PRITS,DT,DEG,DEGL,XMUG,HV,CP)
HQ=1.13*(XKW*3.+DEW*2.*32.2*HV/(XL*XMUG*(TS-TW)))**0.25
HQ=HQ*3600.0
A=DX*DP/(SK*DM)+DP/(HQ*DPO)
HS=1./A
T=T-DT/2.
RETURN
END
$IBFTC PP
SUBROUTINE PP(T,DT,DEG,DEGL,XMUG,HV,CP)
C*****
C*****PHYSICAL PROPERTIES OF WATER VAPOR*****
C*****
T=5.0*(T-32.0)/9.0+273.0
CP=(17.7+.00045*T+.2521E-05*T**2.-.8587E-09*T**3.)/18.0
T=T-273.0
XMUG=(91.9+.00351*T)*.672E-07
DEG=21499.61221632-400.15672932*T+2.5991631*T**2.-.00579602*T**3.
DEG=62.43/DEG
HV=2799.1422976-15.3096*04*T+.20634452*T**2.-.00142704*T**3.
/+0.0000033*T**4.

```



```

HV=HV*453.5924/1054.8
T=9.0*T/5.0+32.0
T1=5.0*((T+DT)-32.0)/9.0
DEG1=21493.61221632-400.15672832*T1+2.5991631*T1**2-.00579602*
/T1**3.
DEG1=62.43/DEG1
RETURN
END

```

£IBFTC MUL

SUBROUTINE MJL(S,T,XMUL)

```

C*****
C ***BAGGASE BLACK LIQUJR VISCOSITY***
C*****
T=5.0*(T-32.)/9.
A=18.276186-.507619*T+.00343*T**2
B=-1.427423+.043054*T-.0003074*T**2
C=.032877-.000898*T+.0000064*T**2
XMUL=EXP(A+B*S+C*S**2)
XMUL=XMUL*.000672
T=9.0*T/5.0+32.
RETURN
END

```

£IBFTC XDEL

SUBROUTINE XDEL(S,T,DEL)

```

C*****
C ***BAGGASE BLACK LIQUJR DENSITY***
C*****
T=5.0*(T-32.)/9.
DEL=1.0207+.0071*S+T*(-.000279-.0000012*S)
DEL=DEL*62.43
T=9.0*T/5.0+32.
RETURN
END

```

£IBFTC XCL

SUBROUTINE XCL(S,T,CL)

```

C*****
C ***BAGGASE BLACK LIQUJR SPECIFIC HEAT***
C*****
T=5.0*(T-32.)/9.
CL=.993-.01258*S+T*(.0000556+.000104*S)
T=9.0*T/5.0+32.
RETURN
END

```

£IBFTC KX

SUBROUTINE KX(S,T,KKL)

```

C*****
C ***BAGGASE BLACK LIQUJR THERMAL CONDUCTIVITY***
C*****
T=5.0*(T-32.)/9.
KKL=.50576-.00268*S+T*(.000899+.00000749*S)
KKL=KKL/1.49
T=9.0*T/5.0+32.
RETURN
END

```

£ENTRY

25. .50 .166

EJOB CGG155,TIME010,PAGES025,NAME R.P.SHLKLA

EIBJOB

EIBFTC MAIN

C*****THIS USES SLIP MODEL FOR BADGER,S DATA SIMULATION

C BADGER,S DATA SIMULATION

COMMON C1,TS,DP

COMMON/LEVEL1/XL

COMMON/LEVEL2/DZ

DO 2 I=1,15

READ 16,XL,DZ,DP

16 FORMAT(3F8.4)

READ 17,WL,TS,T1,S,P

17 FORMAT(5F10.4)

PRINT 3,I

3 FORMAT(* BADGER DATA NO --- *,I3)

WG=0.

PRINT 101

101 FORMAT(*-----INPUT-----*)

PRINT 103,WL,TS,T1,S,P

103 FORMAT(* WL= *,F10.3,* LB-M/SECOND*,/* TS= *,F10.3,* DEGREE F *,/*
/ T=*,F10.2,* DEGREE F *,/* SOLID CONTENT= *,F10.3,* PERCENT *,/*
/= *,F10.3,* LBF/SQ FT *)

PRINT 102

102 FORMAT(*-----OUTPUT-----*)

NE=1

C1=(3.14*DP**2.)/4.

T=T1

CALL EVAP(WL,WG,S,T,P,NE)

2 CONTINUE

STOP

END

EIBFTC EVAP

SUBROUTINE EVAP(WL,WG,S,T,P,NE)

C*****SINGLE EVAPORATOR CALCULATIONS*****

COMMON C1,TS,DP

COMMON/LEVEL1/XL

COMMON/LEVEL2/DZ

XN=XL/DZ

DT=0.0

QT=0.0

N=0

CALL VPT(P,BPT)

BPT=BPT+.19*S*.9/.5.

IF(T-BPT)37,35,35

37 IF(NE.EQ.2) GO TO 35

1 IF(NE.EQ.3) GO TO 35

DZ1=DZ

CALL FHL(WL,T,S,GPRLF,U,Q,UN,DT,DT1,DPT,DZ1,TX,P)

QT1=Q*.4*DP*DZ1

QT=QT+QT1

PRINT 9

P=P+DPT

DPF=-GPRLF*DZ1

PRINT 8,P,DPT,DPF

8 FORMAT(* PRESSURE= *,F 6.8,* LB/SQ FT *,/* TOTAL PRESSURE DROP=

AKA00010

AKA00011

AKA00020

AKA00030

AKA00040

AKA00050

AKA00060

AKA00070

AKA00080

AKA00090

AKA00100

AKA00110

AKA00120

AKA00130

AKA00140

AKA00150

AKA00160

AKA00170

AKA00180

AKA00190

AKA00200

AKA00210

AKA00220

AKA00230

AKA00240

AKA00250

AKA00260

AKA00270

AKA00280

AKA00290

AKA00300

AKA00301

AKA00310

AKA00320

AKA00330

AKA00340

AKA00350

AKA00360

AKA00370

AKA00380

AKA00390

AKA00400

AKA00410

AKA00420

AKA00430

AKA00440

AKA00450

AKA00460

AKA00470

AKA00480

AKA00490

AKA00491

AKA00500


```

/*,F16.8,* LB/SQ FT *,//* FRICTIONAL PRESSURE DROP= *,F16.8,* LB/SQ
/ FT *)
9   FORMAT(110(1F-))
PRINT10,WL,S,T,U,UN,Q,QT1,QT
10  FORMAT(* MASS FLOW RATE OF BLACK LIQUOR= *,F16.4,* LB-M PER SECO
/ND *,//* SOLID CONTENT= *,F16.4,* PERCENT *,//* TEMPERATURE= *,F16.4,*
/.4,* DEGREE FAHRINHEIT *,//* FLOW-VELOCITY= *,F16.4,* FT PER SECONA
/D *,//* OVER-ALL HEAT TRANSFER CO-EFFICIENT= *,F16.4,* BTU/HR.SQ FAKA
/T.DEG F *,//* HEAT FLUX= *,F16.4,* BTU/HR.SQ FT *,//* HEAT TRANSFEAKA
/RRED IN THE SECTION= *,F16.4,* BTU/HR *,//* TOTAL HEAT TRANSFERREDAKA
/ UP TO THIS SECTION= *,F16.4,* BTU/HR *)
N=N+1
PRINT 14,TX
14  FORMAT(* BOILING POINT AT THIS PRESSURE= *,F16.4,* DEGREE F *)
XL=XL-DZ1
PRINT 18,XL,DZ1
18  FORMAT(* REMAINING HEIGHT= *,F16.2,* LENGTH INCREMENT= *,F16.3)
PRINT 13,N
13  FORMAT(* COMPLTATION FOR SECTION *,I3,* HAS BEEN COMPLETED *)
IF(T-TX)2,3,3
2   CONTINUE
PRINT 11
11  FORMAT(* CALCULATION FOR NEXT LENGTH INCREMENT-- *)
DT=0.0
GO TO 1
3   PRINT 15
15  FORMAT(* CALCULATION FOR NONBOILING SECTION IS COMPLETE *)
35  PRINT 36
36  FORMAT(* THERE IS NO NONBOILING SECTION *)
BOILING SECTION NO-1
XN=XL/DZ
N=0
US = 0.
HTB=0.
ST=0.
WT=WL+WG
X=WG/WT
V=WT/(C1*62.43)
TR=0.0
DPI=-40.0
DT=0.
CALL PP(T,DT,DEG,DEGI,XMUG,HV,CP)
DX=400.*.14*DP*CZ*(TS-T)/(3600.*WT*FV)
25  DV=0.
31  DT=0.
DS=0.
CALL EMC(WT,X,S,T,P,V,DX,DS,DT,DPP,DV,TR,Q,WL,WG,U,RGI,RGD)
T=T+DT
X=X+DX
V=V+DV
P=P+DPP
S=S+DS
QT1=Q*.14*DP*DZ
QT=QT+QT1
US=US+U

```

```

HTB=HTB+QT1
PRINT 9
PRINT 28,RGI

PRINT 26,WL,WG,X,T,V,P,S,U,Q,QT1,QT,CPP
28 FORMAT(* VOID FRACTION AT INLET= *,F16.3)
26 FORMAT(* WL= *,F16.4,* LB=M/SECOND *,/* WG= *,F16.4,* LBM/SECOND*,
/* X= *,F16.4,* T= *,F16.4,* DEGREE F *,/* V= *,F16.4,* FT/SECOND*
/* PRESSURE= *,F16.4,* LBF/SQ.FT *,/* SOLID CONTENT= *,F16.4,*
/ PERCENT *,/* OVER ALL HEAT TRANSFER COEFFICIENT= *,F16.4,* BTU/HK
/R.SQFT.DEGREE F *,/* HEAT FLUX= *,F16.4,* BTU/HR.SQFT *,/* HEAT TRA
/ANSFERRED IN THE SECTION= *,F16.4,* BTU/HR *,/* TOTAL HEAT TRANSFE
/RRED= *,F16.4,* BTU/HR *,/* PRESSURE DROP= *,F16.4,* LBF/SQFT *)
PRINT 29,RGO
29 FORMAT(* VOID FRATION AT OUTLET= *,F16.3)
PRINT 9
N=N+1
PRINT 13,N
IF(FLOAT(N)-XN)21,24,24
21 IF((XN-FLOAT(N))-1.)32,33,33
32 DZ=(XN-FLOAT(N))*DZ
33 IF(RGO-0.9) 34,30,30
34 IF(DPP-DP1) 22,23,23
22 IF(RGO-.8)23,30,30
30 TR=3.
PRINT 27
27 FORMAT(* REGION IS ANNULAR *)
23 DP1=DPP
GO TO 25
24 CONTINUE
US=US/FLOAT(N)
US2=FLOAT(N)*HTB/(3.14*DP*XL*ST)
PRINT 41,US,US2
41 FORMAT(* AVERAGE BOILING HEAT TRANSFER COEFFICIENT= *,2F16.2)
PRINT 42 ,HTB
42 FORMAT(* HEAT TRANSFERRED IN THE BOILING SECTION = *,F16.2)
WG=WG*3600.
PRINT 43,WG,P
43 FORMAT(* VAPOR COLLECTED = *,F16.1,* PRESSUR = *,F16.2)
RETURN
END

EIBFTC FML
SUBROUTINE FML(WL,T,S,GPRLF,U,Q,UN,DT,DT1,DPT,DZ,TX,P)
C*****ENERGY,MOMENTUM AND CONTINUITY EQUATIONS FOR SINGLE PHASE LIQUID**
COMMON C1,TS,DP
5 CALL PRLT(WL,T,S,GPRLF,DPF,DZ)
CALL XD(L(S,T,D*CL)
CALL XCL(S,T,CL)
U=WL/(DEL*C1)
DPT=-DEL*DZ-1.0*GPRLF*DZ
DELT=TS-T
CALL XUN(TS,T,S,UN)
Q=UN*DELT
R=DP/2.0
A=Q*2.0-3.14*R*(CZ/(CL*3600.0*DEL*U*C1)

```


B=DZ/(7.7817*100.0*CL)	AKA01490
DT1=A-B	AKA01500
IF(ABS(DT-DT1)-0.1)1,1,2	AKA01510
2 T=T+(DT1-DT)/2.0	AKA01520
DT=DT1	AKA01530
GO TO 5	AKA01540
1 T=T+(DT1+DT)/4.0	AKA01550
P=P+DPT	AKA01560
CALL VPT(P,TX)	AKA01570
P=P-DPT	AKA01580
IF(TX-T)7,6,6	AKA01590
7 IF((T-TX)-.5)6,6,8	AKA01600
8 DZ=DZ*.8	AKA01610
T=T-(DT1+DT)/2.	AKA01620
DT=0.	AKA01630
GO TO 5	AKA01640
6 CONTINUE	AKA01650
RETURN	AKA01660
END	AKA01670
8IBFTC PRLT	AKA01680
SUBROUTINE PRLT(WL,T,S,GPRLF,DPF,DZ)	AKA01690
C*****CALCULATES NONBOILING SECTION LIQUID FRICTIONAL PRESSURE DROP****	AKA01691
COMMON C,TS,DP	AKA01700
CALL MUL(S,T,XMUL)	AKA01710
CALL XDEL(S,T,DEL)	AKA01720
CALL MUL(S,TS,WMUL)	AKA01730
V=WL/(DEL*C1)	AKA01740
RE=DEL*V*DP/XMUL	AKA01750
IF(RE-2100.0)1,1,2	AKA01760
1 F=16.0/RE	AKA01770
F=F*(WMUL/XMUL)**0.38	AKA01780
GO TO 3	AKA01790
2 A=4.0*ALOG10(2.0/(0.00015))	AKA01800
F=(1.0/(A+2.28))**2.0	AKA01810
F=F*(WMUL/XMUL)**0.17	AKA01820
3 DPF=-2.0*F*DZ*V**((2.0)*DEL/(DP*32.2))	AKA01830
GPRLF=-DPF/DZ	AKA01840
RETURN	AKA01850
END	AKA01860
8IBFTC XUN	AKA01870
SUBROUTINE XUN(TS,T,S,UN)	AKA01880
C *****NONBOILING SECTION OVERALL HEAT TRANSFER COEFFICIENT****	AKA01881
CALL XCL(S,T,CL)	AKA01890
CALL MUL(S,T,XMUL)	AKA01900
CALL XDEL(S,T,DEL)	AKA01910
XMUL=3600.0*XMUL	AKA01920
DEL=TS-T	AKA01930
UN=0.06*(CL*DEL*DEL**(.0/XMUL))**0.7	AKA01940
A=1./UN	AKA01950
UN=1./A	AKA01960
RETURN	AKA01970
END	AKA01980
8IBFTC EMC	AKA01990
SUBROUTINE EMC(WT,X,S,T,P,V,DX,DS,DT,DPP,DV,TR,Q,WL,WG,U,RGI,RGO)	AKA02000
C*****SLIP MODEL FOR ENERGY,MOMENTUM AND CONTINUITY EQUATIONS****	AKA02001

	COMMON C1,TS,DP	AKA02010
	COMMON/LEVEL2/DZ	AKA02020
10	CALL PRF(WT,X,T,S,DX,DT,DS,GPRF,RG,P)	AKA02030
	CALL XDEL(S,T,DEL)	AKA02040
	CALL PP(T,DT,DEG,DEGL,XMUG,HV,CP)	AKA02050
	IF(X=0.)22,21,22	AKA02060
21	RGI=0.	AKA02070
	GO TO 23	AKA02080
22	CALL HOLDUP(WT,X,T,S,RLI,RGI,P)	AKA02090
23	DET=(1.-RGI)*DEL+RGI*DEG	AKA02100
	TO=T+DT	AKA02110
	XO=X+DX	AKA02120
	SO=S+DS	AKA02130
	CALL HOLDUP(WT,XC,TO,SO,RLO,RGO,P)	AKA02140
	CALL XDEL(SO,TO,DEL1)	AKA02150
	DET1=(1.-RGO)*DEL1+RGO*DEGL	AKA02160
	DETM=(DET+DET1)/2.	AKA02170
	ACI=(1.-X)**2/(DEL*(1.-RGI))+X**2/(DEG*RGI)	AKA02180
	ACD=(1.-XO)**2/(DEL1*(1.-RGO))+XC**2/(DEGL*RGO)	AKA02190
	AM=ACD-ACI	AKA02200
	DPP=-DETM*DZ-GPRF*DZ-WT**2*AM/(C1**2)	AKA02210
	A=P+DPP	AKA02220
	CALL VPT(A,TX)	AKA02230
	BPE=.19*S*9./5.	AKA02240
	DT=TX+BPE-T	AKA02250
	CALL XCL(S,T,CL)	AKA02260
	AEI=X**3/(DEG**2*RGI**2)+(1.-X)**3/(DEL**2*(1.-RGI)**2)	AKA02270
	AEO=XO**3/(DEGL**2*RGO**2)+(1.-XO)**3/(DEL1**2*(1.-RGO)**2)	AKA02280
	AE=AEO-AEI	AKA02290
	B=WT*3600./(3.14*DP*DZ)	AKA02300
1	C=(1.-X)*CL*DT+HV*DX+X*CP*DT+AE*WT**2/(2.*32.2*778.17*C1**2)+DZ/77	AKA02310
	/8.17	AKA02320
	Q=B*C	AKA02330
	IF(TR.EQ.2.) GC TO 2	AKA02340
	CALL UAFIT(S,DT,DS,Q,RG,U,WT,X,DX)	AKA02350
	GO TO 3	AKA02360
2	CALL UNB(T,S,P,D1,DS,DPP,Q,U,WT,X,DX)	AKA02370
3	Q=U*(TS-T-DT/2.)	AKA02380
	DX1=(Q/B-(1.-X)*CL*DT-X*CP*DT-AE*WT**2/(2.*32.2*778.17*C1**2)-DZ/77	AKA02390
	/78.17)/HV	AKA02400
	IF(ABS(DX-DX1)-.(C03)6,6,5	AKA02410
5	DX=(DX+DX1)/2.	AKA02420
11	FORMAT(60(1H*))	AKA02430
	GO TO 10	AKA02440
6	DX=(DX+DX1)/2.	AKA02450
	WG=WT*X	AKA02460
	WL=WT-WG	AKA02470
	DS=(S*WT*DX)/(WL-WT*DX)	AKA02480
	RETURN	AKA02490
	END	AKA02500
21BFTC	PRF	AKA02510
	SUBROUTINE PRF(WT,X,T,S,DX,DT,DS,GPRF,RG,P)	AKA02520
C*****	CACULATION OF TWO PHASE FRICTIONAL PRESSURE DROP*****	AKA02521
	COMMON C1,TS,DP	AKA02530
	COMMON/LEVEL2/DZ	AKA02540

T=T+DT/2.0	AKA0255C
X=X+DX/2.	AKA0256C
S=S+DS/2.0	AKA0257C
CALL MUL(S,T,XMUL)	AKA0258C
CALL XDEL(S,T,DEL)	AKA0259C
CALL PP(T,DT,DEG,DEG1,XMUG,HV,CP)	AKA0260C
WG=X*WT	AKA0261C
WL=WT-WG	AKA0262C
REG=4.0*WG/(3.14*DP*XMUG)	AKA0263C
REL=4.0*WL/(3.14*DP*XMUL)	AKA0264C
A=WL/WG	AKA0265C
B=DEG/DEL	AKA0266C
C=XMUL/XMUG	AKA0267C
IF(REG-2100.)3,1,1	AKA0268C
1 IF(REL-8000.)3,2,2	AKA0269C
2 VM=(WL/DEL+WG/DEG)/C1	AKA0270C
H=.19*A**.9*B**.5*C**.1*(VM**2./(DP*32.2))**.185	AKA0271C
GO TO 7	AKA0272C
3 IF(REG-2000.)6,4,4	AKA0273C
4 IF(REL-2000.)6,5,5	AKA0274C
5 H=SQRT(A**1.8*B*(**0.2)	AKA0275C
7 PG=EXP(1.54623+0.48518*ALOG(H)+0.0444*(ALOG(H))**2)	AKA0276C
PL=PG/H	AKA0277C
CL=0.046	AKA0278C
XN=0.2	AKA0279C
D=ALOG(H)	AKA0280C
CALL HOLDUP(WT,X,T,S,RL,RG,P)	AKA0281C
ALPA=RL**4.*PL**3.33	AKA0282C
GO TO 14	AKA0283C
6 IF(REG-2000.)10,8,8	AKA0284C
8 IF(REL-2000.)9,10,10	AKA0285C
9 CL=16.0	AKA0286C
CG=0.046	AKA0287C
XN=1.0	AKA0288C
H=SQRT((CL/CG)*REG**(-0.8)*A*B*C)	AKA0289C
D=ALOG(H)	AKA0290C
PG=EXP(1.36397+0.49115*D+0.05209*D**2)	AKA0291C
PL=PG/H	AKA0292C
CALL HOLDUP(WT,X,T,S,RL,RG,P)	AKA0293C
ALPA=RL**2.0*PL**2.0	AKA0294C
GO TO 14	AKA0295C
10 IF(REG-2000.)11,13,13	AKA0296C
11 IF(REL-2000.)13,12,12	AKA0297C
12 CL=0.046	AKA0298C
CG=16.0	AKA0299C
XN=0.2	AKA0300C
H=SQRT((CL/CG)*REL**(0.8)*A*B*C)	AKA0301C
D=ALOG(H)	AKA0302C
PG=EXP(1.35883+0.52153*D+0.05963*D**2-.0013*D**3-.0004*D**4.)	AKA0303C
PL=PG/H	AKA0304C
CALL HOLDUP(WT,X,T,S,RL,RG,P)	AKA0305C
ALPA=RL**4.*PL**3.33	AKA0306C
GO TO 14	AKA0307C
13 H=SQRT(A*B*C)	AKA0308C
CL=16.0	AKA0309C


```

WG=WT*(X+DX/2.)
WL=WT-WG
REL=4.0*WL/(DP*3.14*XMUL)
REG=4.0*WG/(DP*3.14*XMUG)
A=DEL/DEG
B=XMUG/XMUL
C=XKL/DP
H1=C*(1.3+39.*DP)*PRL**C.9*RFL**23*REG**34*A**25*B
D=Q*DP/(HV*XMUL*3600.0)
E=P*DP/ST
F=A-1.0
H2=C*.225*PRL**69*D**79*E**31*F**33
IF (H1-H2)1,1,2
1 H=H2
GO TO 3
2 H=H1
3 CALL XHS(T,DT,H,C,HS)
G=1./H+1./HS
U=1./G
T=T-DT/2.
S=S-DS/2.
P=P-DPP/2.
RETURN
END
$IBFTC UAF
SUBROUTINE UAF(T,S,DT,DS,Q,RG,U,WT,X,DX)
C*****CALCULATES OVERALL HEAT TRANSFER COEFFICIENT FOR ANNULAR FILM FLOW*****
COMMON C1,TS,DP
COMMON/LEVEL2/DZ
T=T+DT/2.
S=S+DS/2.
CALL XCL(S,T,CL)
CALL XDEL(S,T,DEL)
CALL PP(T,DT,DEG,DEG1,XMUG,HV,CP)
WG=WT*(X+DX/2.)
T=(T-32.)*5./9.
ST=75.83076-0.148356*T-.00021*T**2
ST=ST/(453.6*32.2)
T=9.*T/5.+32.
V=WG/(DEG*C1*RG)
A=.012*(V*SQRT(DF*DEG/(ST*32.2)))*.0.5
B=CL*SQRT(DEL*ST*32.2/DP)
H=A*B*3600.0*4.
C*****CALCULATES (STEAM SIDE+WALL) HEAT TRANSFER COEFFICIENT*****
2 FORMAT(* H FOR ANNULAR FLOW=*,F16.4)
CALL XHS(T,DT,H,C,HS)
G=1./H+1./HS
U=1./G
T=T-DT/2.
S=S-DS/2.
RETURN
END
$IBFTC XHS
SUBROUTINE XHS(T,DT,H,Q,HS)
COMMON C1,TS,DP

```

AKA03750
AKA03760
AKA03770
AKA03780
AKA03790
AKA03800
AKA03810
AKA03820
AKA03830
AKA03840
AKA03850
AKA03860
AKA03870
AKA03880
AKA03890
AKA03900
AKA03910
AKA03920
AKA03930
AKA03940
AKA03950
AKA03960
AKA03970
AKA03980
AKA03990
AKA04000
AKA04010
AKA04020
AKA04030
AKA04040
AKA04050
AKA04060
AKA04070
AKA04080
AKA04090
AKA04100
AKA04110
AKA04120
AKA04130
AKA04140
AKA04150

```

COMMON/LEVEL1/XL
COMMON/LEVEL2/DZ
T=T+DT/2.
DX=.0095
DM=.1565
DPO=.166
SK=218.0
R=1./H+DX*DP/(SK*DM)
R=1./R
TW=T+Q/R
TF=TS-3.*(TS-Tw)/4.
XKW=.394/3600.
CALL MUL(S,TF,XMLW)
DEW=.97*62.43
CALL PP(TS,DT,DEG,DEG1,XMUG,HV,CP)
HO=X.13*(XKW**3.*DEW**2.*32.2*HV/(XL*XMUG*(TS-Tw))**.25
HO=HO*3600.0
A=DX*DP/(SK*DM)+CP/(HO*DPO)
HS=1./A
T=T-DT/2.
RETURN
END

```

AKA0416G
AKA0417C
AKA0418C
AKA0419C
AKA0420C
AKA0421C
AKA0422C
AKA0423C
AKA0424C
AKA0425C
AKA0426C
AKA0427C
AKA0428C
AKA0429C
AKA0430C
AKA0431C
AKA0432C
AKA0433C
AKA0434C
AKA0435C
AKA0436C
AKA0437C
AKA0438C
AKA0439C
AKA0440C
AKA0441C
AKA0442C
AKA0443C
AKA0444C
AKA0445C
AKA0446C
AKA0447C
AKA0448C
AKA0449C
AKA0450C
AKA0451C
AKA0452C
AKA0453C
AKA0454C
AKA0455C
AKA0456C
AKA0457C
AKA0458C
AKA0459C
AKA0460C
AKA0461C
AKA0462C
AKA0463C
AKA0464C
AKA0465C
AKA0466C
AKA0467C
AKA0468C

```

SUBROUTINE PP(T,DT,DEG,DEG1,XMUG,HV,CP)
C***PHYSICAL PROPERTIES OF WATER VAPOR***
T=5.0*(T-32.0)/9.0+273.0
CP=(17.7+.00045*T+.2521E-05*T**2.-.85E7E-09*T**3.)/18.0
T=T-273.0
XMUG=(91.9+0.351*T)*.672E-07
DEG=21493.61221632-400.15672832*T+2.5991631*T**2.-.00579602*T**3.
DEG=62.43/DEG
HV=2799.1422976-15.3096504*T+.20634452*T**2.-.00142704*T**3.
/+0.00000353*T**4.
HV=HV*453.5924/1054.8
T=9.0*T/5.0+32.0
T1=5.0*((T+DT)-32.0)/9.0
DEG1=21493.61221632-400.15672832*T1+2.5991631*T1**2.-.00579602*
/T1**3.
DEG1=62.43/DEG1
RETURN
END

```

```

SUBROUTINE KL(S,T,XKL)
C WATER
C***WATER THERMAL CONDUCTIVITY***
T=5.0*(T-32.0)/9.0
XKL=587.*(1.+2.81*.001*(T-20.))
XKL=XKL*.389/4130.
T=9.0*T/5.0+32.0
RETURN
END

```

```

SUBROUTINE MUL(S,T,XMUL)
C WATER
C***WATER VISCOSITY***

```



```

T=5.*(T-32.)/9.
XMUL=1.996531-.0160284*T+.00006745*T**2
XMUL=XMUL*.000672
T=9.*T/5.+32.
RETURN
END

```

```

$IBFTC XDEL
SUBROUTINE XDEL(S,T,DEL)
C WATER
C*****WATER DENSITY****
DEL=.97*62.43
RETURN
END

```

```

$IBFTC XCL
SUBROUTINE XCL(S,T,CL)
C WATER
C*****WATER SPECIFIC HEAT*****
CL=1.0
RETURN
END

```

```

$IBFTC VPT
SUBROUTINE VPT(P,T)
C*****PRESSURE TEMPERATURE EQUATION FOR WATER****
P=P*760./((144.*14.7)
IF(P.LT.30..AND.P.GT.5.) GO TO 2
IF(P.LT.35..AND.P.GT.30.) GO TO 3
IF(P.LT.90..AND.P.GT.35.) GO TO 4
IF(P.LT.150..AND.P.GT.90.) GO TO 5
IF(P.LT.230..AND.P.GT.150.) GO TO 6
IF(P.LT.340..AND.P.GT.230.) GO TO 7
IF(P.LT.520..AND.P.GT.340.) GO TO 8
IF(P.LT.760..AND.P.GT.520.) GO TO 9
IF(P.GT.760.) GO TO 10
2 A=8.184254
B=1791.3
C=238.1
GO TO 1
3 A=8.1393986
B=1767.262
C=236.29
GO TO 1
4 A=8.0886767
B=1739.351
C=234.1
GO TO 1
5 A=8.0464202
B=1715.429
C=232.14
GO TO 1
6 A=8.0136295
B=1695.167
C=230.41
GO TO 1
7 A=7.9845588
B=1678.948

```

```

AKA04680
AKA04690
AKA04700
AKA04710
AKA04720
AKA04730
AKA04740
AKA04750
AKA04760
AKA04761
AKA04770
AKA04780
AKA04790
AKA04800
AKA04810
AKA04820
AKA04821
AKA04830
AKA04840
AKA04850
AKA04860
AKA04870
AKA04871
AKA04880
AKA04890
AKA04900
AKA04910
AKA04920
AKA04930
AKA04940
AKA04950
AKA04960
AKA04970
AKA04980
AKA04990
AKA05000
AKA05010
AKA05020
AKA05030
AKA05040
AKA05050
AKA05060
AKA05070
AKA05080
AKA05090
AKA05100
AKA05110
AKA05120
AKA05130
AKA05140
AKA05150
AKA05160
AKA05170
AKA05180
AKA05190

```

```

C=228.97
GO TO 1
8 A=7.9634288
B=1665.924
C=227.77
GO TO 1
9 A=7.948396
B=1656.39
C=226.86
GO TO 1
10 A=7.9186968
B=1636.909
C=224.92
1 T=(B/(A-ALOG10(P)))-C
T=9.*T/5.+32.
P=P*144.*14.7/760.
RETURN
END

```

```

ENTRY
20. 0.5 0.147 0.0 1230.
0.066 209.75 162.1

```

AKAC5700
 AKAC5710
 AKAC5720
 AKAC5730
 AKAC5740
 AKAC5750
 AKAC5760
 AKAC5770
 AKAC5780
 AKAC5790
 AKAC5800
 AKAC5810
 AKAC5820
 AKAC5830
 AKAC5840
 AKAC5850
 AKAC5860
 AKAC5870

----- INPUT -----

WL= 0.066 LB-M/SECOND
 TS=209.75 DEGREE F
 T= 162.10 DEGREE F
 SOLID CONTENT= 0.000 PERCENT
 P=1230.000 LBF/SQ FT

----- OUTPUT -----

REMAINING HEIGHT= 17.74 LENGTH INCREMENT= 0.256
 CALCULATION FOR NONBOILING SECTION IS COMPLETE
 THERE IS NO NONBOILING SECTION
 AVERAGE BOILING HEAT TRANSFER COEFFICIENT= 470.90 472.41
 HEAT TRANSFERRED IN THE BOILING SECTION= 149427.89
 VAPOR COLLECTED= 151.7 PRESSURE= 1230.00

£JOB CGG153,TIME008,PAGES020,NAME A.K.AGARWAL

£IBJOB

£IB6TC MAIN

C*****
C*****THIS USES SLIP MODEL FOR BADGER,S DATA SIMULATION
C*****THIS PROGRAMME USES BOELTER EQUATION FOR NONBOILING HEAT TRANSFER
C*****

COMMON C1,TS,DP
COMMON/LEVEL1/XL
COMMON/LEVEL2/DZ
READ 16,XL,DZ,DP
READ 17,WL,TS,TI,S,P

16 FFORMAT(3F8.4)

17 66FORMAT(5F10.3)

WG=0.

PRINT 110

110 FORMAT(1H1)

P9INT 101

1 66FORMAT&----- INPUT-----*)

P9INT 03,WL,TS,TI,S,P

1 3 66FORMAT(* WL= *,F10.3,* LB-M/SECOND*,/* TS= *,F10.3,* DEGREE F *,/*
/ T=*,F10.2,* DEGREE F *,/* SOLID CONTENT= *,F10.3,* PERCENT *,/* P
/= *,F10.3,* LBF/SQ FT *)

PRINT 102

102 FORMAT(*----- OUTPUT-----*)

NE=1

C1=(3.14*DP**2.)/4.

T=TI

PRINT 1

1 FORMAT(*-----EVAPORATOR NO 1-----*)

LL ESAP(WL,WG,S,T,P,NE)

STOP

554

£IB6TC EVAP

C*****&&&&*****
C*****SINGLE EVAPORATOR CALCULATIONS*****
C*****

SUBROUTINE EVAP(WL,WG,S,T,P,NE)

COMMON C1,TS,DP

COMMON LEVEL1/XL

COMMON/LEVEL2/DZ

K=2

XN=XL/DZ

DT=0.0

QT=0.0

N=0

3A3L V7T(P,BPT)

BPT=BPT+.19*S*.9./S.

IF(T-BPT)37,35,35

37 IF(NE.EQ.2) GO TO 35

IF(NE.EQ.3) GO TO 35

1 DZ1=DZ

CALL EML(WL,T,S,GPRLF,U,Q,UN,DT,DT1,DPT,DZ1,TX,P)

QT1=Q*3.14*DP*DZ1

QT=QT+QT1

PRINT 9

```

P=P+DPT
DPF=-G7RLF*DZ1
PRINT 8,P,DPT,DPE
8  FORMAT(* PRESSURE= *,F16.8,* LB/SQ FT *,/** TOTAL PRESSURE DROP=
/*=F16.8,* LB/SQ FT *,/** FRICTIONAL PRESSURE DROP= *,F16.8,* LB/SQ
/ FT *)
9  60FORMAT(110(1H-))
P9INT 10,WL,S,T,U,UN,Q,QT1,QT
10  FORMAT(* MASS FLOW RATE OF BLACK LIQUOR= *,F16.4,* LB-M PER SECO
/N4 *,/** SOLID CONTENT= *,F16.4,* PERCENT *,/** TEMPERATURE= *,F16
/.4,* DEGREE FAHRENHEIT *,/** FLOW-VELOCITY= *,F16.4,* FT PER SECON
/4 &,/** OVER-ALL HEAT TRANSFER CO-EFFICIENT= *,F16.4,* BTU/HR.SQ F
/T.DEG F *,/** HEAT FLUX= *,F16.4,* BTU/HR.SQ FT *,/** HEAT TRANSFER
/RRD IN THE SECTION= *,F16.4,* BTU/HR *,/** TOTAL HEAT TRANSFERRED
/ UP TO THIS SECTION= *,F16.4,* BTU/HR *)
N=N+1
799NT 14, TX
14  FORMAT(* BOILING POINT AT THIS PRESSURE= *,F16.4,* DEGREE F *)
XL=XL-DZ1
P9INT 8,XL,DZ1
18  FORMAT(* REMAINING HEIGHT= *,F16.2,* LENGTH INCREMENT= *,F16.3)
PRINT 13,N
13  FORMAT(* COMPUTATION FOR SECTION *,13,* HAS BEEN COMPLETED *)
PRINT 9
IF(N.LT.K) GO TO 51
IF(N.EQ.K) GO TO 52
52  PRINT 53
53  FORMAT(1H1)
K=K+2
51  CONTINUE
IF(T-TX)2,3,3
2  CONTINUE
79INT 1
11  FORMAT(* CALCULATION FOR NEXT LENGTH INCREMENT-- *)
DT=0.0
GO TO 1
3  79INT 15
1  6RMA0(* CALCULATION FOR NONBOILING SECTION IS COMPLETE *)
35  P9INT 36
36  66RMA0(* THERE IS NO NONBOILING SECTION *)
BOILING SECTION NO-1
K=3
X5=XL/+Z
N=0
US = 0.
HTB=0.
ST=0.
WT=WL+WG
X=WG/WT
V=WT/(C1*62.43)
TR=2.0
DP1=-40.0
DO=0.
3+3L PP(T,DT,D G,DESI,XMUG,HV,CP)
DX=400.*3.14*DP*DZ*(TS-T)/(3600.*WT*HV)
25  DV=0.

```



```

31 DT=0.
   DS=0.
   CALL EMC(WT,X,S,T,P,V,DX,DS,DT,DPP,4V,TR,Q,WL,WG,U,RGI,RGO)
   T=T+DT
   X=X+DX
   V=V+DV
   P=P+DPP
   S=S+DS
   CALL XDEL(S,T,DEL)
   V=WL/(DEL*C1*(1.-RGO))
   QT1=Q*3.14*DP*4Z
   QT=QT+QT1
   US=US+U
   HTB=HTB+QT1
   ST=ST+TS-T+DT/2.
   PRINT 9
   PRINT 28,RGI
28  FORMAT(* VOID FRACTION AT INLET= *,F16.3)
   PRINT 26,WL,WG,X,T,V,P,S,U,Q,QT1,QT,DPP
26  FORMAT(* WL= *,F16.4,* LB-M/SECOND *,/* WG= *,F16.4,* LBM/SECOND*,
  /* X= *,F16.4,* T= *,F16.4,* DEGREE F *,/* V= *,F16.4,* FT/SECOND
  / *,/* PRESSURE= *,F15.4,* LBF/SQ.FT *,/* SOLID CONTENT= *,F16.4,*
  / PERCENT *,/* OVER ALL HEAT TRANSFER CO-EFFICIENT= *,F16.4,* BTU/H
  /R.SQFT.DEGREE F *,/* HEAT FLUX= *,F16.4,* BTU/HR.SQFT *,/* HEAT TR
  /ANSFERRED IN THE SECTION= *,F16.4,* BTU/HR *,/* TOTAL HEAT TRANSFER
  /RRED= *,F16.4,* BTU/HR *,/* PRESSURE DROP= *,F16.4,* LBF/SQFT *)
   PRINT 29,RGO
29  FORMAT(* VOID FRACTION AT OUTLET= *,F16.3)
   PRINT 9
   N=N+1
   PRINT 13,N
   IF(N.LT.K) GO TO 54
   IF(N.EQ.K) GO TO 55
55  PRINT 53
   K=K+3
54  CONTINUE
   IF(FLOAT(N)-XN)21,24,24
21  IF((XN-FLOAT(N))-1.)32,33,33
32  DZ=(XN-FLOAT(N))*DZ
33  IF(RGO-0.9) 34,30,30
34  IF(DPP-DP1) 22,23,23
22  IF(RGO-.8)23,30,30
30  TR=3.
   PRINT 27
27  FORMAT(* REGION IS ANNULAR *)
23  DP1=DPP
   GO TO 25
24  CONTINUE
   US=US/FLOAT(N)
   US2=FLOAT(N)*HTB/(3.14*DP*XL*ST)
   PRINT 41,US,JS2
41  FORMAT(* AVERAGE BOILING HEAT TRANSFER COEFFICIENT= *,2F15.2)
   PRINT 42,HTB
42  FORMAT(* HEAT TRANSFERRED IN THE BOILING SECTION = *,F15.2)
   RETURN
   END

```


£IBFTC EML

C*****
C*****ENERGY, MOMENTUM AND CONTINUITY EQUATIONS FOR SINGLE PHASE LIQUID***
C*****

```
      SUBROUTINE EML(WL,T,S,GPRLF,U,Q,UN,DT,DT1,DPT,DZ,TX,P)
      COMMON C1,TS,DP
5     CALL PRLT(WL,T,S,GPRLF,DPF,DZ)
      CALL XDEL(S,T,DEL)
      CALL XCL(S,T,CL)
      U=WL/(DEL*C1)
      DPT=-DEL*DZ-1.0*GPRLF*DZ
      DELT=TS-T
      CALLXUN(WL,T,S,DT,UN)
      Q=UN*DELT
      R=DP/2.0
      A=Q*2.0*3.14*R*DZ/(CL*3600.0*DEL*U*C1)
      B=DZ/(7.7817*100.0*CL)
      DT1=A-B
      IF(ABS(DT-DT1)-0.1)1,1,2
2     T=T+(DT1-DT)/2.0
      DT=DT1
      GO TO 5
1     T=T+(DT1+DT)/4.0
      P=P+DPT
      CALL VPT(P,TX)
      P=P-DPT
      IF(TX-T)7,6,6
7     IF((T-TX)-.5)6,6,8
8     DZ=DZ*.8
      T=T-(DT1+DT)/2.
      DT=0.
      GO TO 5
6     CONTINUE
      RETURN
      END
```

£IBFTC PRLT

C*****
C*****CALCULATES NONBOILING SECTION LIQUID FRICTIONAL PRESSURE DROP***
C*****

```
      SUBROUTINE PRLT(WL,T,S,GPRLF,DPF,DZ)
      COMMON C1,TS,DP
      CALL MUL(S,T,XMUL)
      CALL XDEL(S,T,DEL)
      CALL MUL(S,TSYWMUL)
      V=WL/(DEL*C1)
      RE=DEL*V*DP/XMUL
      IF(RE-2100.0)1,1,2
1     F=16.0/RE
      F=F*(WMUL/XMUL)**0.38
      GO TO 3
2     A=4.0*ALOG10(2.0/0.00015)
      F=(1.0/(A+2.28))**2.0
      F=F*(WMUL/XMUL)**0.17
3     DPF=-2.0*F*DZ*V**2.0*DEL/(DP*32.2)
      GPRLF=-DPF/DZ
      RETURN
```

END

IBFTC XUN

C *****NONBOILING SECTION OVERALL HEAT TRANSFER COEFFICIENT*****
C *****

SUBROUTINE XJN(WL,T,S,DT,UN)

COMMON C1,TS,DP

CALL XCL(S,T,CL)

CALL MUL(S,T,XMUL)

CALL XDEL(S,T,DEL)

CALL KL(S,T,XKL)

XL=3.2

TF=123.03

H1=250.

TF=5.*(TF-32.)/9.

TW=TS-.05*TS

5 TW=5.*(TW-32.)/9.

BETA=(.0001612*(TW-TF)+.0000031*(TW**2-TF**2))*5./((.9962245+.0001
/612*TF**2+.0000031*TF**2)*(TW-TF)*9.)

TF=9.*TF/5.+32.

TW=9.*TW/5.+32.

10 XNU=H1*DP/XKL

GZM=3600.*WL*CL/(XKL*XL)

A=3.14*XNU/GZM

B=ALOG((2.+A)/(2.-A))

F1=A/B

IF(A.LT.1.85) GO TO 1

IF(A.LT.2.) GO TO 2

F2=0.

GO TO 3

1 F2=1.-.408*A

GO TO 3

2 F2=-1.63*(A-2.)

3 C=32.2*DP**4*BETA*(TW-TF)*3600.*DEL**2*CL/(XMUL*XKL*XL)

XNU=1.75*F1*(GZM+.0722*F2*C**-.84)**.33

H2=XNU*XKL/DP

IF(ABS(H1-H2).LT.5.) GO TO 4

H1=H2

GO TO 10

4 H=(H1+H2)/2.

CALL XHS(T,DT,H,Q,HS)

G=1./H+1./HS

UN=1./G

RETURN

END

IBFTC EMC

C *****SLIP MODEL FOR ENERGY,MOMENTUM AND CONTINUITY EQUATIONS*****
C *****

SUBROUTINE EMC(WT,X,S,T,P,V,DX,DS,DT,DPP,DV,TR,Q,WL,WG,U,RSI,RGD)

COMMON C1,TS,DP

COMMON/LEVEL2/DZ

10 CALL PRF(WT,X,T,S,DX,DT,DS,GPRF,RG,P)

CALL XDEL(S,T,DEL)

CALL PP(T,DT,DEG,DEG1,XMUG,HV,CP)

IF(X-0.) 2,21,22


```

21 RGI=0.
   GO TO 23
22 CALL HOLDUP(WT,X,T,S,RLI,RGI,P)
23 DET=(1.-RGI)*DEL+RGI*DEG
   TO=T+DT
   XO=X+DX
   SO=S+DS
   CALL HOLDUP(WT,XO,TO,SO,RLO,RGO,P)
   CALL XDEL(SO,TO,DEL1)
   DET1=(1.-RGO)*DEL1+RGO*DEG1
   DETM=(DET+DET1)/2.
   ACI=(1.-X)**2/(DEL*(1.-RGI))+X**2/(DEG*RGI)
   ACO=(1.-XO)**2/(DEL1*(1.-RGO))+XO**2/(DEG1*RGO)
   AM=ACO-ACI
   DPP=-DETM*DZ-GPRF*DZ-WT**2*AM/(C1**2)
   A=P+DPP
   CALL VPT(A,TX)
   BPE=.19*S*9./5.
   DT=TX+BPE-T
   CALL XCL(S,T,CL)
   AE1=X**3/(DEG**2*RGI**2)+(1.-X)**3/(DEL**2*(1.-RGI)**2)
   AEO=XO**3/(DEG1**2*RGO**2)+(1.-XO)**3/(DEL1**2*(1.-RGO)**2)
   AE=AEO-AE1
   B=WT*3600./(3.14*DP*DZ)
1  C=(1.-X)*CL*DT+HV*DX+X*CP*DT+AE*WT**2/(2.*32.2*778.17*C1**2)+DZ/77
   /8.17
   Q=B*C
   IF(TR.5Q.2.) GO TO 2
   CALL UAF(T,S,DT,DS,Q,RG,U,WT,X,DX)
   GO TO 3
2  CALL UNBIT(S,P,DT,DS,DPP,Q,U,WT,X,DX)
3  Q=U*(TS-T-DT/2.)
   DX1=(Q/8-(1.-X)*CL*DT-X*CP*DT-AE*WT**2/(2.*32.2*778.17*C1**2)-DZ/77
   /78.17)/8V
   IF(ABS(DX-DX1)-.0003)6,6,5
5  DX=(DX+DX1)/2.
11 FORMAT(60(1H*))
   GO TO 10
6  DX=(DX+DX1)/2.
   WG=WT*X
   WL=WT-WG
   DS=(S*WT*DX)/(WL-WT*DX)
   RETURN
   END

```

EIBFTC PRF

```

C*****
C**8**3ACULATION OF TWO PHASE FRICTIONAL PRESSURE DROP*****
C*****

```

```

   SUBROUTINE PRF(WT,X,T,S,DX,DT,DS,GPRF,RG,P)
   COMMON C1,TS,DP
   COMMON/LEVEL2/DZ
   T=T+DT/2.0
   X=X+DX/2.
   S=S+DS/2.0
   CALL MUL(S,T,XMUL)
   CALL XDEL(S,T,DEL)

```

```

CALL PP(T,DT,DEG,DEGI,XMUG,HV,CP)
WG=X*WT
WL=WT-WG
REG=4.0*WG/(3.14*DP*(MUG)
REL=4.0*WL/(3.14*DP*(XMUL)
A=WL/W7
B=DEG/DEL
C=XMUL/XMUG
IF(REG-2100.)3,1,1
1 IF(REL-8000.)3,2,2
2 VM=(WL/DEL+WG/DEG)/C1
H=.19*A**.9*3**.5*C**.1*(VM**2./(DP*32.2))**.185
GO TO 7
3 IF(REG-2000.)6,4,4
4 IF(REL-2000.)6,5,5
5 H=SQRT(1**1.8*B*C**0.28
7 PG=EXP(1.54623+0.48518*ALOG(H)+0.0444*(ALOG(H))**2)
PL=PG/H
CL=0.046
XN=0.2
D=ALOG(H)
CALL HOLDUP(WT,X,T,S,RL,RG,P)
ALPA=RL**4.*PL**3.33
GO TO 14
6 IF(REG-2000.)10,8,8
8 IF(REL-2000.)9,10,10
9 CL=16.0
CG=0.046
XN=1.0
H=SQRT((CL/CG)*REG**(-.8)*A*B*C)
D=ALOG(8)
PG=EXP(1.36397+0.49115*D+0.05209*D**2)
PL=PG/H
CALL HO-DUP(WT,X,T,S,RL,RG,P)
ALPA=RL**2.0*PL**2.0
GO TO 14
10 IF(REG-2000.)11,13,13
11 IF(REL-2000.)13,12,12
12 CL=0.046
CG=16.0
XN=0.2
H=SQRT((CL/CG)*REL**(0.8)*A*B*C)
D=ALOG(H)
PG=EXP(1.35883-0.52153*D+0.05963*D**2-.0013*D**3-.0004*D**4.)
PL=PG/H
CALL HOLDUP(WT,X,T,S,RL,RG,P)
ALPA=RL**4.*PL**3.33
GO TO 14
13 H=SQRT(A*B*C)
CL=16.0
XN=1.0
D=ALOG(H)
PG=EXP(1.0705+.5520*D+.0881*D**2-.0026*D**3-.0012*D**4)
PL=PG/H
CALL HOLDUP(WT,X,T,S,RL,RG,P)
ALPA=RL**2.0*PL**2.0

```



```

14 E=SQRT(ALPHA/RL)
F=2.*(4.*WL/3.14)**(2.*XN)*CL*XMUL**XN/(DP**(5.-XN)*DEL*32.2)
G=ALPHA**(XN-2.)*F**(5.-XN)
GPRF=F*G
PRD=GPRF*DZ
T=T-DT/2.
X=X-DX/2.
S=S-DS/2.
RETURN
END

```

£IBFTC HOLDUP

```

C*****
C*****CACULATION OF VOID FRACTION IN THE TUBE*****
C**&&&&*****&*****

```

SUBROUTINE HOLDUP(WT,X,T,S,RL,RG,P)

COMMON C1,TS,DP

CALL XDEL(S,T,DEL)

DT=0.

CALL PP(T,DT,DEG,DEG1,XMUG,HV,CP)

A=-2.*(DEG*(1.-X)**2-DEL*X**2)

B=DEG*(1.-X)**2-4.*DEL*X**2

C=2.*D+L*X**2

D=SQRT(B**2-4.*A*C)

R7=(-2+D)/(2.*+)

RG1=(-B-D)/(2.*A)

IF(RG-1.)4,5,5

5 RG=RG1

GO TO 3

4 IF(RG-0.)2,3,3

2 RG=RG1

3 RL=1.-RG

RETURN

END

£IBFTC UNB

```

C**&&&&*****&*****
C*****CACULATES OVERALL HEAT TRANSFER COEFFICIENT FOR NUCLEAT BOILING***
C*****

```

SUBROUTINE UNB(T,S,P,DT,DS,DPP,Q,U,WT,X,DX)

COMMON C1,TS,DP

COMMON/LEVEL2/DZ

T=T+DT/2.

S=S+DS/2.

P=P+DPP/2.

CALL MUL(S,T,XMUL)

CALL XDEL(S,T,DEL)

CALL XCL(S,T,CL)

CALL KL(S,T,XKL)

CALL PP(T,DT,DEG,DEG1,XMUG,HV,CP)

T=(T-3.)*5./9.

ST=75.83076-0.148336*T-.00021*T**2

ST=ST/(1488.6*32.2)

T=9.*T/5.+32.

793=CL*XMUL*3600./XKL

WG=WT*(X+DX/2.)

WL=WT+WG

REL=4.-0*WL/(DP*3.14*XMUL)

```

REG=4.0*WG/(DP*3.14*XMUG)
A=DEL/4+G
B=XMUG/XMUL
C=XKL/DP
H1=C*(1.3+39.*DP)*PRL**0.9*REL**.23*REG**.34*A**.25*B
D=Q*DP/(HV*XMUL*3600.0)
E=P*DP/ST
F=A-1.0
82=C*.25*PRL**.69*D**.69*F*.31*F**.33
IF (H1-H2) 1,1,2
1 H=H2
GO TO 3
2 H=H1
3 CALL XHS(T,DT,8,Q,HS)
G=1./H+1./HS
U=1./G
T=T-DT/2.
S=S-DS/2.
P=P-D7P/2.
RETURN
END

```

£IBFTC UAF

C*****
C*****CALCULATES OVERALL HEAT TRANSFER COEFFICIENT FOR ANNULAR FILM FLOW**
C**&***&***&***&*****

SUBROUTINE UAF(T,S,DT,DS,Q,RG,U,WT,X,DX)

COMMON C1,TS,DP

C64MON/LEVEL2/DZ

T=T+DT/2.

S=S+DS/2.

CALL XCL(S,T,CL)

CALL XDEL(S,T,DEL)

CALL PP(T,DT,DEG,DEG1,XMUG,HV,CP)

WG=WT*(X+DX/2.)

T=(T-32.)*5./9.

ST=75.83076-0.148356*T-.00021*T**2

ST=ST/(453.6*32.2)

T=9.*T/5.+32.

5=W7/(4EG*C1*RG)

A=.0128(V*SQRT(DP*DEG/(ST*32.2)))**0.5

B=CL*SQRT(DEL*ST*32.2/DP)

H8=B*3600.0*4H

2 FORMAT(* H FOR ANNULAR FLOW= *,F15.4)

CALL XHS(T,DT,H,Q,HS)

G=1./H+1./HS

U=1./G

T=T-DT/2.

S=S-DS/2.

RETURN

END

£IB6TC 78S

C**&*****
C*****CALCULATES (STEAM SIDE+WALL) HEAT TRANSFER COEFFICIENT*****
C*****

S SUBROUTINE XHS(T,DT,H,Q,HS)

COMMON C1,TS,DP

REI
END

£IBFTC MUL

```
C*****
C**&***WATER VISCOSITY****
C*****&&***&*****
      SUBROUTINE MUL(S,T,XMUL)
      T=5.*(T-32.)/9.
      XMUL=1.996531-.0160284*T+.00006745*T**2
      XMUL=XMUL*.000672
      T=9.*T/5.+32.
      RETURN
      END
```

£IBFTC XDEL

```
C*****
C*****WATER DENSITY****
C*****
      SUBROUTINE XDEL(S,T,DEL)
      DEL=.97*62.43
      RETURN
      END
```

£IBFTC XCL

```
C**&***&*****
C*****WATER SPECIFIC HEAT*****
C*****
      SUBROUTINE XCL(S,T,CL)
      CL=1.0
      RETURN
      END
```

£IBFTC VPT

```
C*****
C*****PRESSURE-TEMPERATURE EQUATION FOR WATER****
C*****
C&***&&***&*****8*****
      SUBROUTINE VPT(P,T)
      P=P*760./((144.*14.7)
      IF(P.LT.30..AND.P.GT.5.) GO TO 2
      IF(P.LT.35..AND.P.GT.30.) GO TO 3
      IF(P.LT.90..AND.P.GT.35.) GO TO 4
      IF(P.LT.150..AND.P.GT.90.) GO TO 5
      IF(P.LT.230..AND.P.GT.150.) GO TO 6
      IF(P.LT.340..AND.P.GT.230.) GO TO 7
      IF(P.LT.520..AND.P.GT.340.) GO TO 8
      IF(P.LT.760..AND.P.GT.520.) GO TO 9
      IF(P.GT.760.) GO TO 10
2      A=8.184254
      B=1791.3
      C=238.1
      GO TO 1
3      A=8.139 986
      B=1767.262
      C=236.29
      GO TO 1
4      A=8.0886767
      B=1739.351
      C=234.1
      GO TO 1
5      A=8.0464 02
```



```

      B=1715.429
      C=232.14
      GO TO 1
6     A=8.0116295
      B=1695.157
      C=230.41
      GO TO 1
7     A=7.9845588
      B=1678.948
      C=228.97
      GO TO 1
8     A=7.9634288
      B=1665.924
      C=227.77
      GO TO 1
9     A=7.948396
      B=1656.39
      C=226.86
      GO TO 1
10    A=7.9186968
      B=1636.909
      C=224.92
1     T=(B/(A-ALOG10(P)))-C
      T=9.*T/5.+32.
      P=P*144.*14.7/760.
      RETURN
      END

```

ENTRY

0.	0.5	0.147		
-068	178.93	127.03	0.0	1075.